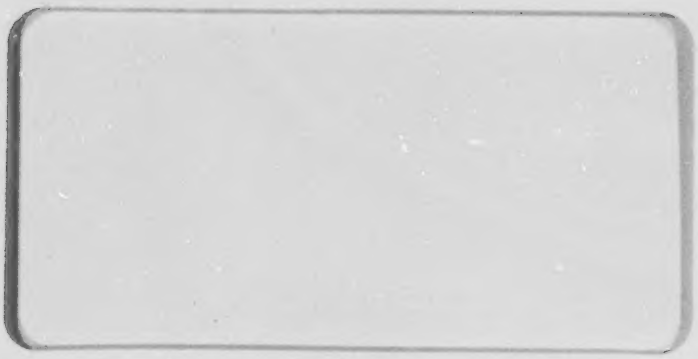

ADM

Acres Davy McKee

ACRES DAVY McKEE ENGINEERING INC.
Burlington, Ontario

FERRIC CHLORIDE LEACHING OF SULPHIDE MINERAL CONCENTRATES
PHASE I
FINAL REPORT

DSS FILE NO. 23SQ.23440-3-9083
CONTRACT SERIAL NO. OSQ83-00172
ADM CONTRACT NO. C1016.00

164 pages

April 12, 1984

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JUL 30 1984

ADM
Acres Davy McKee

Engineers and Managers

April 12, 1984
C1016.00

Mr. B. Lucas
Head, Leaching Section
Extractive Metallurgy Lab/CANMET
Department of Energy, Mines and Resources
555 Booth Street
Ottawa, ON
K1A 0G1

Dear Sir:

DSS File No. 23SQ.23440-3-9083
Contract Serial No. 0SQ83-00172
ADM Contract No. C1016.00

Please find enclosed 20 copies of the final report entitled "Ferric Chloride Leaching of Sulphide Mineral Concentrates - Phase I".

We have endeavored to clarify all the points raised during your review of the draft version of the report and we trust that the final version meets with your approval. We are willing to discuss the contents of the document at your convenience.

Also enclosed is one copy of the documentation pertaining to the literature search, performed as part of this study.

Should you have any questions, please do not hesitate to contact us. It has been a pleasure working with you on this interesting project.

Yours very truly,

David J. King

D.J. King, P.Eng.
Project Manager

Enclosure

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TITLE: Ferric Chloride Leaching of Sulphide Mineral Concentrates - Phase I

FINAL REPORT

By

Acres Davy McKee Engineering Inc.

April, 1984

Work on this project was conducted under the auspices of the Canada Centre for Mineral and Energy Technology, Energy, Mines and Resources, Canada

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CANMET SCIENTIFIC AUTHORITY: Mr. B. Lucas
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Résumé a l'attention de la direction

Résumé

(a) D'après les renseignements recueillis, suite au dépouillement des publications relatives à ce sujet, le schéma du procédé de lessivage au chlorure ferrique en deux passes, tel que proposé par CANMET, a subi des modifications afin de produire les fondements pour le design d'une usine-pilote de 10 kg/h.

(b) Le design du système utilise par l'usine a atteint l'étape ou les diagrammes du procédé de lessivage, la liste de l'équipement requis et le plan montrant la disposition de cet équipement ont été préparés.

(c) L'usine-pilote est unique en ce sens que l'argent, le plomb, le zinc et le cuivre y sont extraits en un seul et même procédé intégré. Plusieurs autres usines ont réussi à extraire individuellement des métaux ou certaines combinaisons de metaux, mais aucune n'a tenté l'extraction de la combinaison argent/plomb/zinc/cuivre.

(d) Les critères établis pour le design de la plupart des sections de l'usine sont bases sur de solides données recueillies auprès de CANMET ou dans les publications dépouillées. Des essais effectués en laboratoire seront cependant nécessaires à la phase II du programme pour certaines sections de l'usine.

(e) La plupart de l'équipement utilisé par l'usine est facilement disponible. Les matériaux de construction pouvant résister à ce milieu extrêmement corrosif sont généralement faciles à trouver, bien qu'en fin de compte l'opération de l'usine-pilote sera nécessaire pour attester de leur conformité.

(f) La comparaison, sur les plans économique et technique, entre le procédé de lessivage au chlorure ferrique utilisé par CANMET, la technologie de l'acide sulfurique sous pression de Sherritt Gordon

et le procédé de grillage-sulfuration et lessivage employé par le NBRPC ne pourra être entreprise qu'après l'obtention de données suffisantes, lesquelles seront recueillies lors de l'opération de l'usine. Optimiser les conditions du procédé en question, en tant que partie intégrante de l'operation de l'usine-pilote, s'avère nécessaire avant que ce procédé soit considéré pour toute utilisation commerciale.

(g) Etabli à partir de devis fournis par les fabriquants d'équipement et de paramètres internes basés sur le côut total de l'équipement, l'investissement en capital pour la construction de l'usine se chiffrerait à 764 800$ (±25%). Ce devis se base sur certaines suppositions décrites à la Section 5 de ce rapport.

(h) Une vérification de l'équipement disponible et approprîe a l'usine-pilote a montré qu'une réduction de 65 800$ sur le coût d'investissement global pourrait être effectuée si l'usine était construite dans les installations de l'ORF.

1 - EXECUTIVE SUMMARY, CONCLUSIONS AND RECOMMENDATIONS

1 - EXECUTIVE SUMMARY, CONCLUSIONS AND RECOMMENDATIONS

1.1 - Executive Summary

(a) The two-stage ferric chloride leaching flowsheet proposed by CANMET has been modified, using the information gathered from the literature search, to produce a design basis for mini-plant to treat 10 kg/h concentrate.

(b) The process design for the mini-plant has been developed to a stage where process flow diagrams, an equipment list and a preliminary layout have been prepared.

(c) The mini-plant is unique in that silver, lead, zinc and copper are extracted in one integrated flowsheet. Several other plants have successfully extracted individual metals, or combinations of metals, but none have been used to extract the silver/lead/zinc/copper combination.

(d) The design criteria developed for most of the mini-plant sections are based upon well-established data obtained from CANMET or the literature search. Bench-scale testing, to be carried out during Phase II of the program, will be required in order to more clearly define the design criteria for certain sections of the mini-plant.

(e) Most of the equipment contained in the mini-plant is available off-the-shelf. Materials of construction suitable for the extremely corrosive environment are readily available, although operation of the mini-plant is necessary to ultimately confirm their suitability.

(f) Technical and economic comparisons between the CANMET ferric chloride leach process, Sherritt Gordon's pressure sulphuric acid technology and NBRPC's sulphation roast leach process can only be undertaken when sufficient data from the mini-plant test regime has been obtained.

Optimization of process conditions, as part of the mini-plant test regime, is required before a commercial scale operation can be considered.

(g) Based upon quotations from equipment suppliers and employing in-house factors based on total equipment cost, an order-of-magnitude capital cost for the mini-plant has been estimated at $764,800.

This capital cost is based upon certain assumptions which are outlined in section 5 of this report.

(h) An audit of equipment available and suitable for use in the pilot plant has determined that a capital cost reduction of $65,800 would be realized if the mini-plant were constructed at ORF's facilities.

1.2 - Conclusions

(a) The literature search did not provide a relevant reference in which silver, lead, zinc and copper can be recovered in a single flowsheet based upon ferric chloride leaching of complex base metal sulphide concentrates. Information, however, was available on the extractions of individual metals, and other combinations of metals, from the ferric chloride medium.

(b) The assessment of CANMET test work revealed that it had progressed to a stage where a potential flowsheet had been identified. However, in this flowsheet only certain sections had been tested sufficiently to enable the determination of adequate design criteria.

(c) A preferred flowsheet has been developed based upon the CANMET work, supplemented by information obtained from the literature search. The flowsheet includes the following

- two-stage leaching using ferric chloride
- recovery of silver by copper cementation
- recovery of lead, prior to zinc and copper recovery, by crystallization of lead chloride
- recovery of lead by fused-salt electrolysis of lead chloride
- recovery of zinc by solvent extraction with TBP and aqueous chloride electrowinning
- recovery of copper by solvent extraction with Acorga P5300 and copper sulphate crystallization
- oxidation of recycle lixiavants using chlorine gas recovered from zinc electrowinning.

(d) Bench-scale test work to confirm the design criteria for silver cementation and copper solvent extraction is desirable.

(e) Design criteria have been developed and a preliminary design for a mini-plant has been produced.

Continuous pilot scale test work is necessary to confirm

- equipment specification and sizing
- materials of construction
- product quality and recoveries
- water balance
- effluent quality
- process control.

1.3 - Recommendations for Future Work

In order to advance the assessment of the CANMET ferric chloride leach process, it is recommended that Phase II of the program be implemented. Phase II of the program will include the following

(a) Bench scale test work to confirm silver cementation, copper solvent extraction design criteria and the suitability of adding calcium chloride to the leach for sulphate control.

(b) Pilot plant test work to confirm the technical feasibility of the integrated flowsheet developed in Phase I of this study.

(c) Supplementary bench-scale test work, using real solutions, in conjunction with pilot plant testing.

(d) Detailed technical and economic evaluation of the CANMET ferric chloride leach process, using the pilot plant test results in comparison with the New Brunswick Research and Productivity Council's sulphation-roast leach process and Sherritt Gordon's pressure sulphuric acid technology.

2 - STATEMENT OF WORK

2 - STATEMENT OF WORK

2.1 - Introduction

CANMET is currently evaluating a number of hydrometallurgical processes as alternatives to the conventional pyrometallurgical treatment of sulphide ores in order to improve the recovery of metals and reduce the impact on the environment. The hydrometallurgical processes under consideration are

(a) pressure sulphuric acid (PSA) process of Sherritt-Gordon

(b) sulphation roast-leach (SRL) process of the N.B. Research and Productivity Council

(c) ferric chloride leaching

(d) dry-way chlorination.

The evaluation of the first two processes is well advanced, but ferric chloride leaching and dry-way chlorination are not nearly as well developed.

In order to allow ferric chloride leaching to be compared with the PSA and SRL processes, CANMET has decided to accelerate the development of this process by constructing and operating a continuous mini-plant to provide substantive data for process evaluation.

This program has been divided into two phases and the execution of Phase I was awarded to Acres Davy McKee Engineering Inc. acting as the main contractor with the Ontario Research Foundation acting as subcontractor. The objective of Phase I is to fully assess the process of ferric chloride leaching of complex sulphide concentrates with respect to its feasibility. The objective has been accomplished by the following steps

(a) Compilation of all known literature relating to the work, through the use of computer and other searches of the open literature.

(b) Discussion and assessment of the work conducted at CANMET.

(c) Assessment of the relative merits of proposed unit operations based on best current technological practices with respect to

- Applicability to differing complex sulphide ore types.
- Metal extraction.
- Subsequent metal recovery and impurity levels.
- Regeneration of lixiviant and subsequent recycle to leach.
- Waste management and treatment of process effluents.
- Process control.
- Materials of construction.

(d) Development of a process design and a capital cost estimate for a continuous mini-plant to conduct continuous closed circuit operation of ferric chloride leaching of sulphide ores in order to implement the further testwork and to confirm process data for the unit operations already gathered by CANMET. The plant has been designed to allow maximum flexibility in operating conditions and to utilize existing equipment, whenever possible.

(e) From the above work, conclusions have been drawn on the feasibility of the ferric chloride process and recommendations have been made regarding a future work plan.

The detailed scope of work of the project is contained in Section 2.2 of this study.

2.2 - Scope of Work

The objectives of Phase I of the project have been outlined in Section 2.1. The following represents the work that has been undertaken by Acres Davy McKee Engineering Inc. and the Ontario Research Foundation in order to satisfy the objectives.

2.2.1 - Literature Search

A comprehensive computer based literature search on the technical feasibility of ferric chloride as well as chloride assisted leaching has been conducted.

The detailed literature search included the following aspects.

(a) The Characteristics of Complex Sulphide Ores

The characteristics of the New Brunswick ores have been compared against other complex sulphide ores (Caribou-Heathsteel Canadian), McArthur - Australian, Spanish, Norwegian and Kuroko - Japanese). The mineralogical information gathered regarding liberation size has provided useful information on size reduction, leaching and liquid/solid separation.

(b) Process Related Factors

The literature search included evaluation and recovery of elements such as Cu, Zn, Fe, Ag and Pb from chloride assisted systems in general and ferric chloride leach systems in particular. The chemistry of extraction and separation of halometallic complexes arising from chloride assisted base metal aqueous systems can be very different from sulphuric assisted extraction systems. Hence, the literature search has

critically evaluated such differences while designing the flowsheet. Relevant details such as impurity content, product quality, product recovery (e.g. Ag, S·), material handling, and materials engineering from chloride assisted extraction systems have been critically evaluated.

(c) Waste Management

The literature search has assessed the quality of effluents and tailings produced from chloride assisted systems for environmental disposal and storage (e.g. Jarosite).

2.2.2 - Assessment of CANMET Work

After completion of the literature search and evaluation of the state of the art in ferric chloride leaching technology, a detailed assessment of the results obtained from the CANMET test program has been undertaken. The technical feasibility of the overall ferric chloride leaching process has been evaluated based upon the literature search and the CANMET data. A work plan has been developed in order that necessary design data, which may have been omitted from both the published literature and the CANMET program, may be obtained.

The scope of the evaluation of the CANMET results included the following areas

- extraction efficiency of base metals
- recovery of base metals
- control of impurities
- physico-chemical character of final products and their compatibility to market specifications
- solution balance in closed cycle
- leachant regeneration
- by-product (e.g. Ag, S^0) recovery

. liquid/solid separation
. bleed stream control
. waste management
. materials handling
. materials of construction.

2.2.3 - Mini-Plant Design

In order to enable the confirmation of the design data obtained during the CANMET program on a pilot plant scale, the plant has been designed with a high degree of inherent flexibility. Optimum flowrates, temperatures, pressures, residence times are just some of the process variables which will require determination before a comprehensive design for a commercial facility can be undertaken. Careful consideration has been given to incorporating monitoring devices in the design in order to maximize the effectiveness of the plant as a research tool. The layout of the plant has been another important aspect since its labour intensive nature requires that ease of access to all equipment be given prime consideration. Materials of construction have been specified to withstand the most severe operating conditions which may be encountered.

The design of the mini-plant has been carried out using well established and proven design procedures in order to obtain a reliable, safe and operable plant at minimum cost. The process design has involved the production of the following documentation.

(a) Process Flow Diagram (PFD) showing main equipment and process lines.
(b) Equipment list containing sizes, types and materials of construction.
(c) Conceptual plot plan/general arrangement drawing.
(d) Process description and waste disposal options.

The process design work carried out under Phase I of the project will allow detailed design work (process and instrumentation diagrams, electrical and instrumentation drawings, etc.) to be performed under Phase II of the project.

2.2.4 - Capital Cost Estimate

Using a combination of vendor pricing and in-house data on equipment, labour and transportation costs in combination with cost factors prevalent in the construction of pilot plants, an order-of-magnitude capital cost estimate for the mini-plant has been developed. The capital cost estimate includes

- equipment costs
- equipment installation
- civil and structural
- building modifications
- electrical
- piping
- insulation and painting
- instrumentation
- services.

Attention has been given to ways in which to minimize costs (e.g. utilizing existing buildings, offices and equipment) and these options have been outlined in the report.

3 - INFORMATION RETRIEVAL AND EVALUATION

3 - INFORMATION RETRIEVAL AND EVALUATION

3.1 - Literature Search

3.1.1 - General

A computer assisted review was undertaken on the literature pertaining to ferric chloride and chlorides assisted leaching systems. Key articles were collected and secondary references were then consulted. References used appear in the appendix.

(a) Strategy for Evaluation of Process Data

Since one of the objectives of the program is to fully assess the process of ferric chloride leaching of complex sulphide concentrates and to enable its comparison with other processes, emphasis was made in the literature review on whether key data generated at the bench scale level are suitable for scale up. These data must define the key operating parameters for each unit operation in the process. The experimentation must have been conducted at a sufficient scale to permit a reasonable level of correlation with the operational characteristics of a pilot plant. The data should also have been gathered on a representative test solution.

An important factor in this study has been the decision of what constituted "sufficient" technical data to allow the testing of the pilot scale process.

The final flowsheet has been developed as a result of a systematic evaluation of the process proposed by CANMET, using the principles outlined above.

3.1.2 - Background

The discovery of massive ore bodies of readily separable minerals in deposits of economic grade is becoming rare. The best deposits have already been exploited and reserves of such ores are insufficient to meet future needs. Attention is now being turned to the considerable reserves which exist in complex sulphide ore deposits.

In Canada, the Brunswick deposit is well known, both for its abundance and difficulty of mineral separation. Relatively large deposits of other similar complex deposits are found elsewhere in the world. These include (i) McArthur (Australia), (ii) Kuroko (Japan), (iii) Sulitjelma (Norway), (iv) La Zarza Tharsis (Spain) (1) and (v) Lahanos (Turkey).

The mean metal content of these deposits is in the range 0.3-3% Cu, 0.3-3% Pb, 0.2-10% Zn, 30-100 g/t Ag and a negligible amount of Au. However, due to their complexity and to the lack of suitable economic metals recovery processes, these reserves are in general, being used mainly as raw materials for sulphuric acid production. Interest is now being shown in developing methods which would be able to economically recover the metals values from these ores.

The mineral structure in complex sulphide ores can vary significantly. Chalcopyrite and galena may show intimate crystal growth whereas sphalerite is generally coarse enough to be easily liberated. Chalcopyrite can also be disseminated in pyrites of micrometer size crystals. In some ores, the complex sulphides are interbedded in a matrix of fine-grained chalcopyrite and sphalerite. Boe et al (2) have plotted the complexity of various metal sulphides and their mineral structure, based upon liberation size (figure 3.1.1). The graph contains three separate sections.

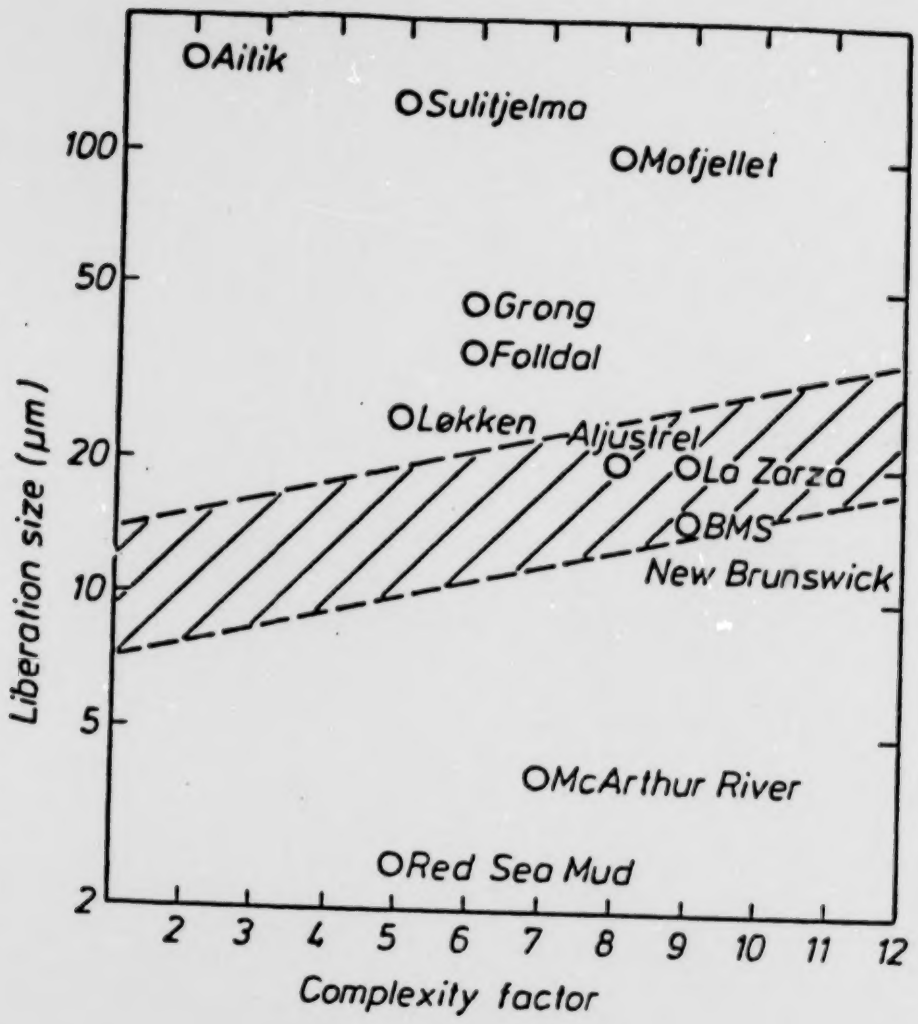


FIGURE 3.1.1

Sulfide Ores Plotted According to Complexity and Liberation Size

(a) Ores in the upper part of the graph, for example Sulitjema, are relatively easy to handle compared to those in the lower part.

Currently, smelter complexes are successfully being used to treat these ores. Side streams or contaminated products from one section are transferred to another part of the same refining complex to complete the separation. However, since such refinery complexes are extremely expensive, large ore deposits are required to justify the expenditure. Where the ore deposits are limited, hydrometallurgical processes (see (b) below) may be more appropriate.

(b) Ores in the lower part of the graph, for example New Brunswick or McArthur River, are extremely complex and finely grained. These ores are characterized by the fine intergrowth of lead, copper and zinc. Flotation methods of separation are usually unsatisfactory since liberation of finely intergrown materials requires very fine grinding. Such grinding is not only expensive but leads to high mineral losses through oxidation and slimes in tailings.

A further complication is that fine grinding alone will not necessarily liberate one mineral from the others, since the minerals do not fracture regularly along grain boundaries. Thus most of the flotation concentrates are characterized by poor recovery of the desirable material and high levels of cross-contamination of one mineral by another.

Economical exploitation of these ores will depend upon the utilization of suitable concentration methods to produce bulk concentrates which can then be treated in order to recover the metal values.

Since conventional pyrometallurgical techniques cannot readily handle such bulk concentrates, other processes, essentially hydrometallurgical, are being developed. These processes are the subject of the literature search.

The ideal hydrometallurgical treatment process would

- maximize overall metal recovery
- minimize processing costs
- minimize emissions of toxic metals.

The hydrometallurgical processes have a further advantage over pyrometallurgical operations in that sulphur dioxide emissions are substantially reduced.

(c) Ores in the banded area, i.e. the middle of the graph, for example La Zarza, could be treated by either pyrometallurgical or hydrometallurgical means. The choice of process will depend upon several factors such as the extent of the deposit and environmental regulations.

3.1.3 - Hydrometallurgical Processes

Hydrometallurgical methods which have been studied for sulphide ores or concentrates include

(a) Acid Sulphate

i) Bacterial
ii) Ferric sulphate
iii) Copper sulphate
iv) Roasting - water or acid leaching
v) Sulphuric acid with oxygen or air
vi) Sulphuric acid with acid production from in-situ pyrite oxidation
vii) Sulphuric acid under reducing conditions.

(b) Chloride Systems

i) Ferric chloride
ii) Ferric chloride - copper chloride
iii) Hydrochloric acid
iv) Roasting - sodium chloride leaching
v) Chlorine

(c) Nitric Acid

(d) Alkaline Systems

i) Ammonia - ammonium sulphate
ii) Sodium hydroxide

The scope of the present program involves the survey and assessment of chloride systems, particularly ferric chlorides, as the leachant for complex sulphide ores/concentrates. The following flowsheets are hence identified and assessed. An assessment of the flowsheet developed by CANMET for complex metal sulphides is discussed in Section 3.3.

i) Broken Hill Associates, Australia (BHAS Process)
ii) CLEAR (Duval) Process
iii) Cymet (Cypress Metallurgical Process)
iv) Dextec (Dextec Metallurgical PTY Ltd. Process)
v) ELKEM (Elkem A/S Norway) Process
vi) ELKEM/ICI (ICI/ES Process)
vii) MINIMET RECHERCHE (I Metal Group Process)
viii) Sherritt-Gordon-Cominco Process
ix) U.S.B.M. Lead (U.S. Bureau of Mines Process)
x) SOX (Selective Oxidation Process)
xi) CSIRO (Commonwealth Scientific and Industrial Research Organization Australian Process)
xii) U.S.B.M. Copper Process.

3.1.4 - Copper-Recovery Processes

The BHAS, CLEAR, CYMET, U.S.B.M., COMINCO-SHERRITT GORDON and SOX processes are essentially developed for chalcopyrite concentrates. Their application for mixed sulphide concentrates is not known. Copper is recovered by either electrowinning or hydrogen reduction. Details of the processes are discussed under the appropriate sections.

i) BHAS Process (3)

In this process a copper sulphide by-product from lead smelting is subjected to chloride/sulphate leaching at 85°C and atmospheric pressure. Copper and zinc are separated by solvent extraction steps and electrowon. Lead and silver remain with the tails, due to the low chloride level in the lixiviant. Pilot plant test work using chalcopyrite and chalcocite concentrate has been conducted on a scale equivalent to a 5 ton per day copper plant.

The process is simple, and produces copper free of silver and has very few recycle streams. However, the process does not at present show silver and lead recovery steps.

ii) CLEAR (Duval) Process (4,5)

This process (see figure 3.1.2) is a combination of cupric-ferric chloride and is applied principally to treat chalcopyrite concentrate.

No details to recover Zn, Pb, Ag have been disclosed.

The copper in the first stage leach liquor is 85 - 90 percent cuprous. This value is increased to practically 100 percent cuprous by reaction with copper prior to electrowinning.

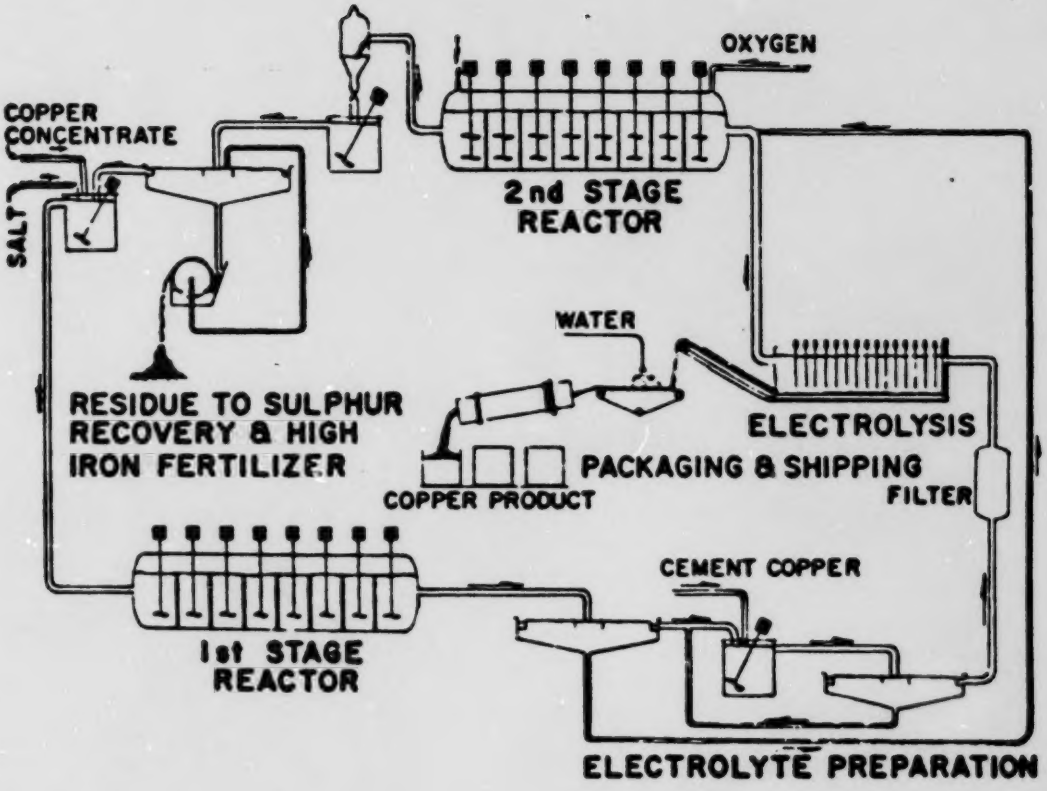


FIGURE 3.1.2
CLEAR (Duval) Process
Rev. 1

Spent electrolyte is the initial lixiviant for the second stage leach. The second stage reactions produce spent solid residue and regenerated leach solution for recycle to the first stage. Copper is recovered as granules in diaphragm cells. Oxygen is required for the second stage reaction at 150°C to oxidize the remaining cuprous ions and regenerate ferric ions to the desired level. The iron leached from the feed concentrate is rejected as potassium jarosite together with beta-ferric hydroxide.

The plant operated under commercial conditions from 1976 - 1982. The plant is currently shut down for modification due to the depressed copper market. The modifications will include the installation of new electrolytic cells with reduced cathode-anode spacing, enlarged electrode area and improved electrolyte agitators. Prevention of silver contamination of copper is also being investigated.

Process Advantages

1. Low energy required for electrowinning copper from cuprous solution.
2. Cuprous and ferrous ions are oxidized in the anolyte compartment. Since chlorine is not liberated a collection/ recycling system is not required.
3. Cuprous ion can be converted to cupric, which is a strong oxidant in high chloride concentrations.
4. Elemental sulphur is a reaction product and does not present a serious disposal problem.

Disadvantages

1. No mention of ability to treat complex sulphides. Process may be specific to copper and not readily adaptable to complex ores.
2. Copper product is a coarse powder which would either be sold at below premium prices, or require additional refining.
3. High salt concentration in solution is required to maintain cuprous solubility.

4. Anolyte agitation is required, which complicates the electrolytic cell design. Agitation improves cuprous ion contact with the anode.
5. Second stage leach requires mild autoclave conditions, 138°C and 40 psig, to effect oxidation and complete copper dissolution. Pure oxygen is also used which adds to the cost.

iii) Cymet Process (6,10)

This process (see figure 3.1.3) is a combination of cupric-ferric leach with cupric ions used as the first stage lixiviant and ferric ions for the second stage. The process is applied only to clean chalcopyrite concentrate.

The high chloride concentrations are designed to produce high cuprous ion concentrations which may be separated as cuprous chloride crystals. Iron is rejected as sodium or hydronium jarosite. Metallic copper is produced by reducing the cuprous chloride crystals with hydrogen in a fluid bed reactor. The HCl produced is returned to the leach circuit. The high level of sulphate produced in the first stage of leach results in practically all of the iron being rejected as jarosite. Control of iron precipitation and maintaining specific iron concentrations in the circuit discharge are critical factors. It is not clear whether the process can compete against a smelter for treating clean concentrate. The copper agglomerates produced in the fluid bed reactor are impure and must be fired or electro-refined to meet wire bar grade specifications. All of the silver in the test reactor reported to copper.

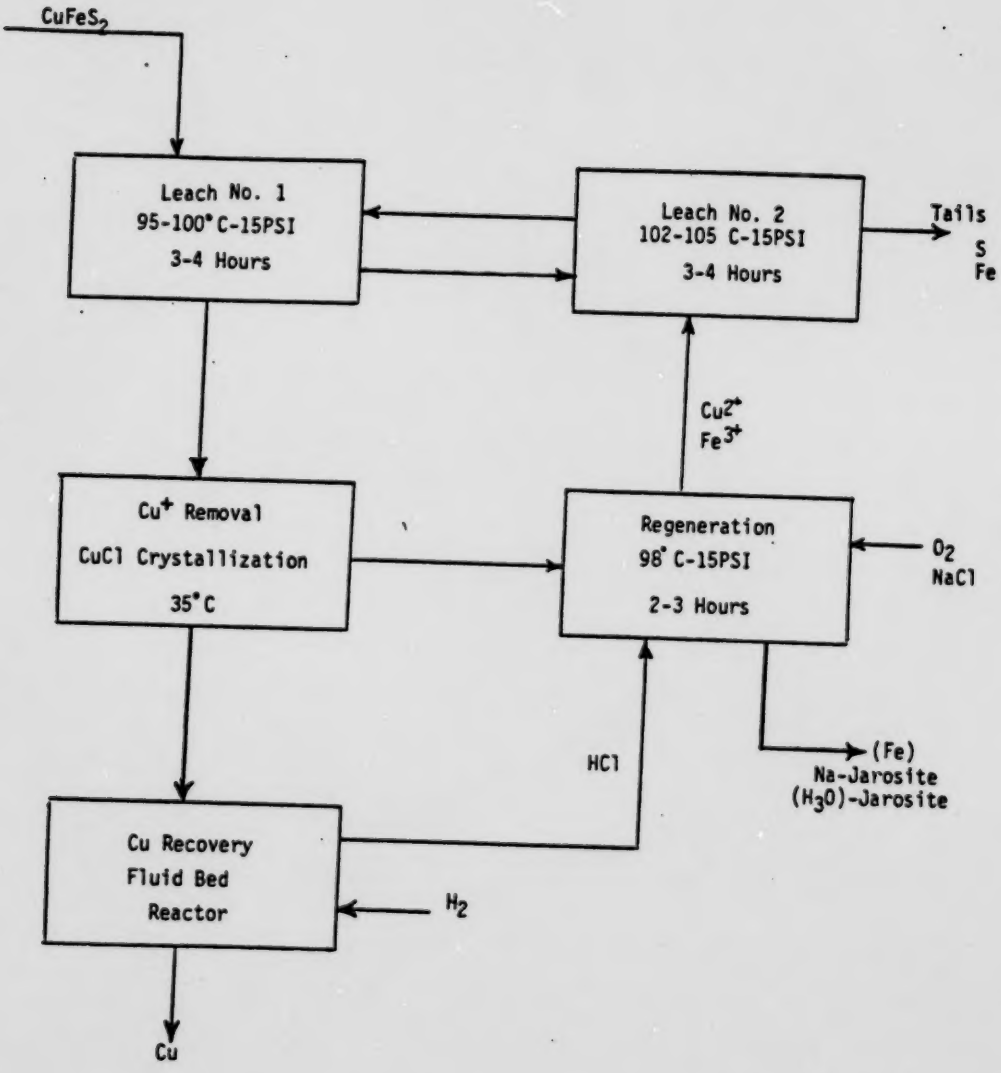


FIGURE 3.1.3

Cymet Process
Rev. 1

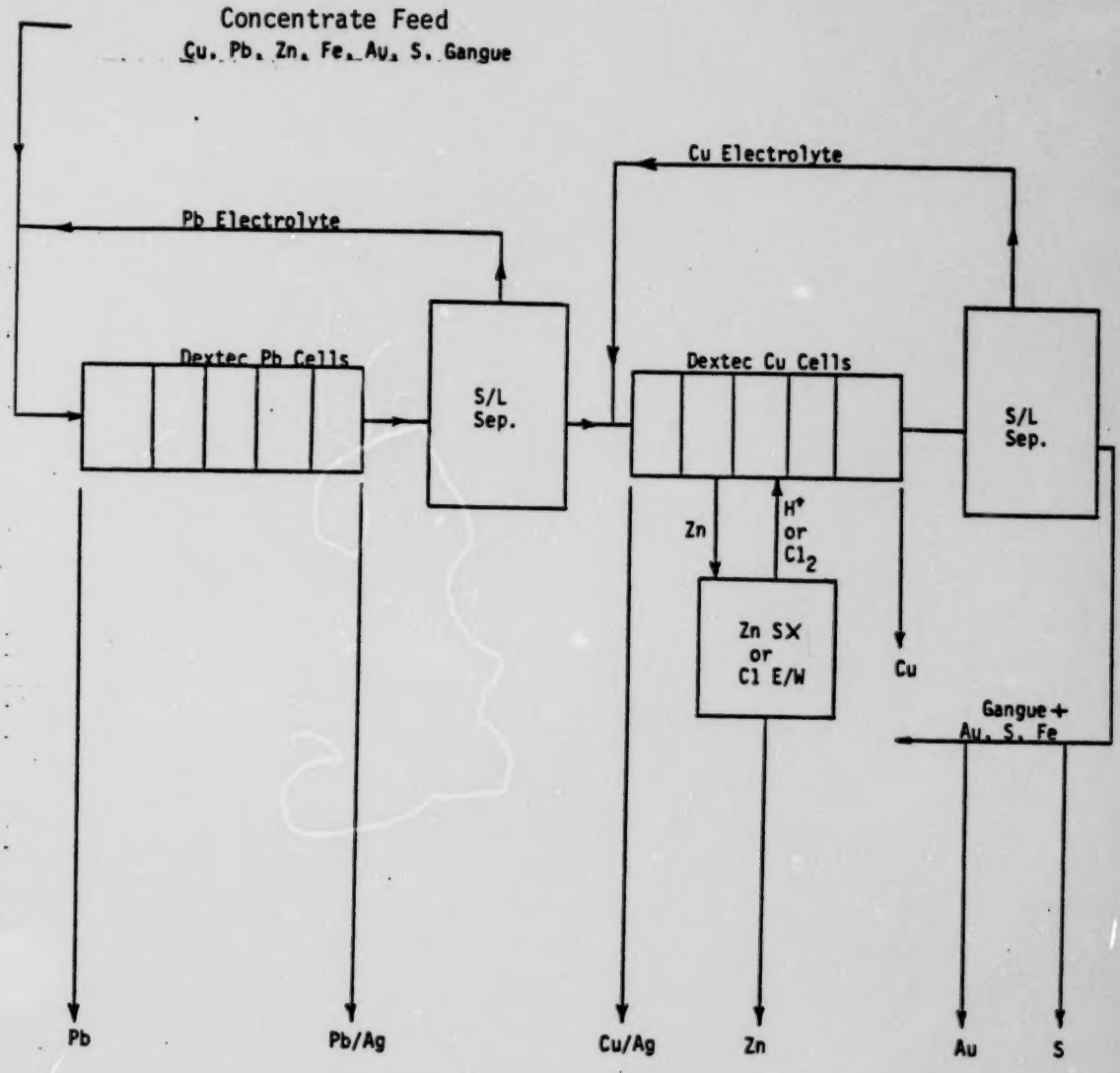


FIGURE 3.1.4

Conceptual Dextec Flow Sheet for Mixed Sulphides

iv) Dextec Process (7)

The Dextec Process (see figure 3.1.4) involves anodic dissolution of copper from chalcopyrite concentrate in brine solutions in the presence of oxygen. The process is carried out in a diaphragm cell and copper is deposited at the cathode. Copper product contains 1,000 ppm of impurities. The fine residue contains elemental sulphur, iron oxides and gold. Silver reports to the copper during electrolysis. By adopting zinc solvent extraction, the process may find application in treating Cu/Pb/Zn concentrates on large or small scales. The process may be configured to extract impurities such as lead from copper concentrates.

The operation of the diaphragm cell to treat slurries in chloride systems needs to be proven. Details on recycle streams, residue handling and disposal need to be assessed.

v) Sherritt-Cominco Process (3,10, US patent 4,097,271)

This process (Figure 3.1.7) is a combination cupric-ferric, high chloride leach process. The process is more complicated and sophisticated than most other chloride processes. The process is one of the few requiring reagents other than hydrogen, oxygen and chlorine; namely, butadiene and caustic.

The flowsheet has a leach circuit in which part of the cupric and ferric ions are generated in-situ by injection of chlorine. The leach circuit operates at 15 psi and 95 - 105°C with a retention time of about 3 - 5 hours. Cuprous ion is removed from the leach liquor by injecting butadiene gas and cooling to about 20°C which permits reduction of cuprous concentration to 5 g/L. The cuprous chloride complex is decomposed by slurrying and heating to 65°C in a stirred reactor in a partial vacuum.

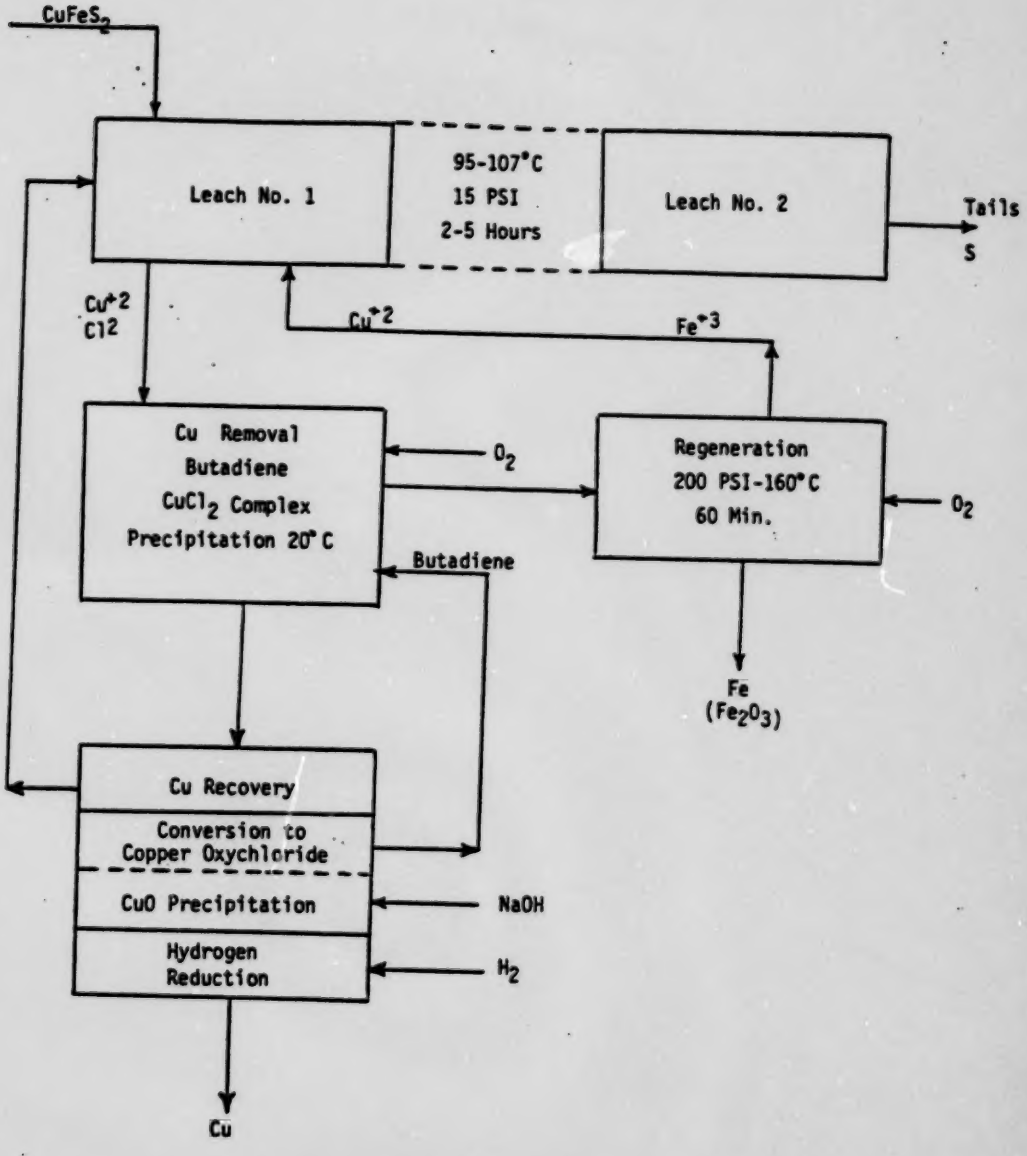


FIGURE 3.1.7

Sherritt-Cominco Process

Rev. 1

Butadiene is recovered as a gas for recycle. In the regeneration circuit, the liquor from the cuprous chloride complex precipitation is processed in a pressure reactor at 200 psi and 160°C. Oxygen precipitates ferric oxides, while regenerating cupric and ferric ions. The regenerated liquor is recycled to the leach circuit. Cuprous chloride is converted to metal in the following three stages

i) oxidizing it to copper oxychloride with air or oxygen
ii) hydrolyzing the oxychloride to CuO with NaOH
iii) reducing the CuO with hydrogen to produce high purity copper containing no silver.

The advantage of this route includes

i) liquid-solid separation problems between the stages in two stage leach circuits are removed
ii) copper produced does not require additional refining
iii) the process reagents are substantially recovered
iv) the circuit capacity is higher than comparable flowsheets due to the higher copper concentrations of 40 g/l.

The disadvantages of the process could include its complexity and the economics of treating dirty concentrates.

vi) SOX Process (3)

The process involves a cupric chloride leach of copper concentrates with the recovery of copper by solvent extraction and electrowinning. Copper is recovered from sulphate solutions in a refined form. Silver is recovered from the chloride stream by a separate circuit. Pilot scale studies using 1 - 3 week campaigns have been conducted.

Detailed process data and economics are not available.

3.1.5 - Total Metals Recovery Processes

The ELKEM, ELKEM/ICI, MINEMET processes feature the treatment of total chloride systems. All component metals are recovered. Details of the processes are discussed under the appropriate sections to follow. The quality of copper produced by this group of methods is high compared to the group in Section 3.1.4 except for Cominco-Sherritt.

i) ELKEM Process (8)

The process (see figures 3.1.5, 3.1.5A) features the treatment of complex metal sulphide concentrates in total chloride systems with the recovery of all valuable metals. The process includes counter-current leaching using $FeCl_3/CuCl_2$. Lead is recovered by crystallization of lead chloride. Zinc is separated by solvent extraction using TBP, and recovered by electrowinning in diaphragm cells. Copper is recovered by purging of iron as akagniete (FeOOH) and carphosiderite $(Fe_3(SO_4)_2(OH)_5 \cdot H_2O)$. Cu, Zn, and Pb are recovered at efficiencies of 98 - 99 percent and Ag at 95 percent.

The process can handle dirty complex sulphide concentrates unlike many of the other processes where development is restricted to pure or dirty chalcopyrite concentrates. The economics of the process are not attractive for copper recovery, due to silver reporting to the copper product. The overall economics of the process are not clear. The flowsheet is complicated.

ii) ELKEM/ICI Process (3)

This process is similar to the Elkem Process, except for copper recovery. The direct electrowinning is replaced by solvent extraction and fluidized bed electrolysis. Copper is extracted with ACORGA AS 5443. Copper contained in the pregnant strip solution is recovered by fluidized bed electrolysis. In order to

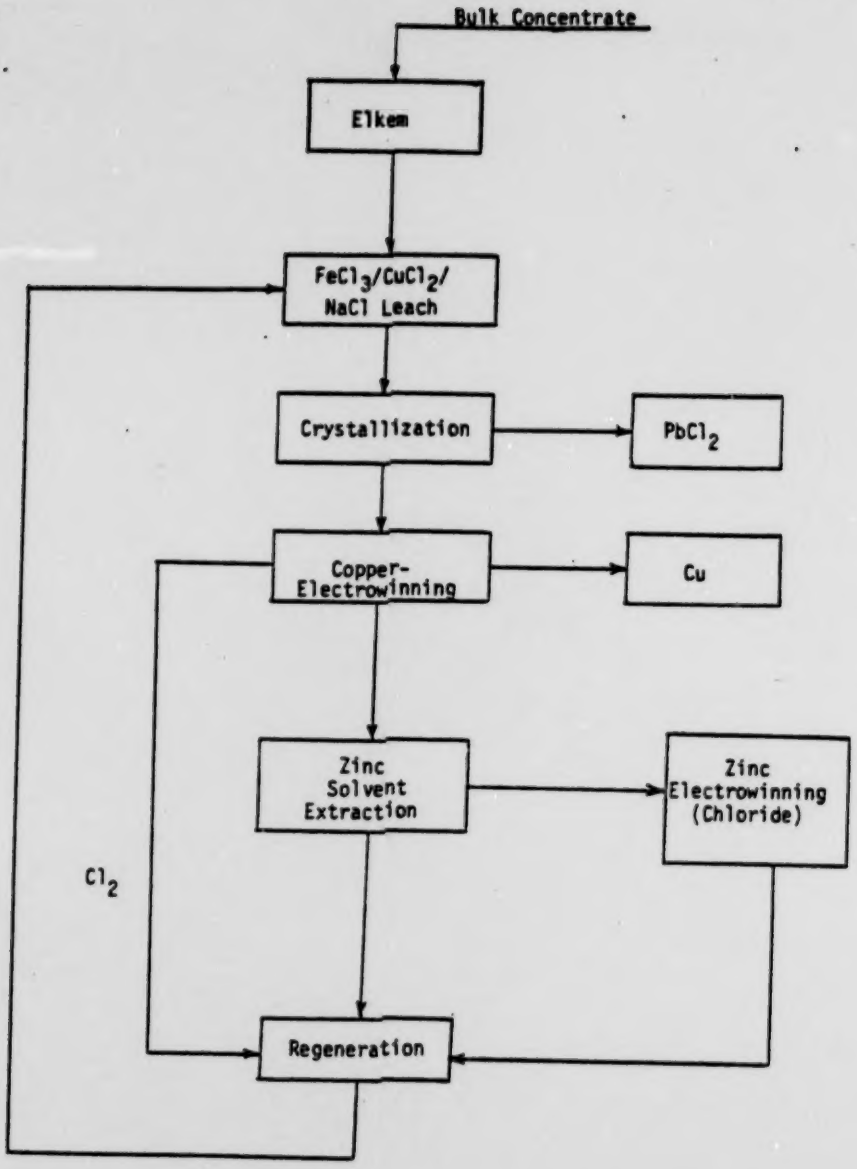


FIGURE 3.1.5

Outline Circuit of the Elkem Process

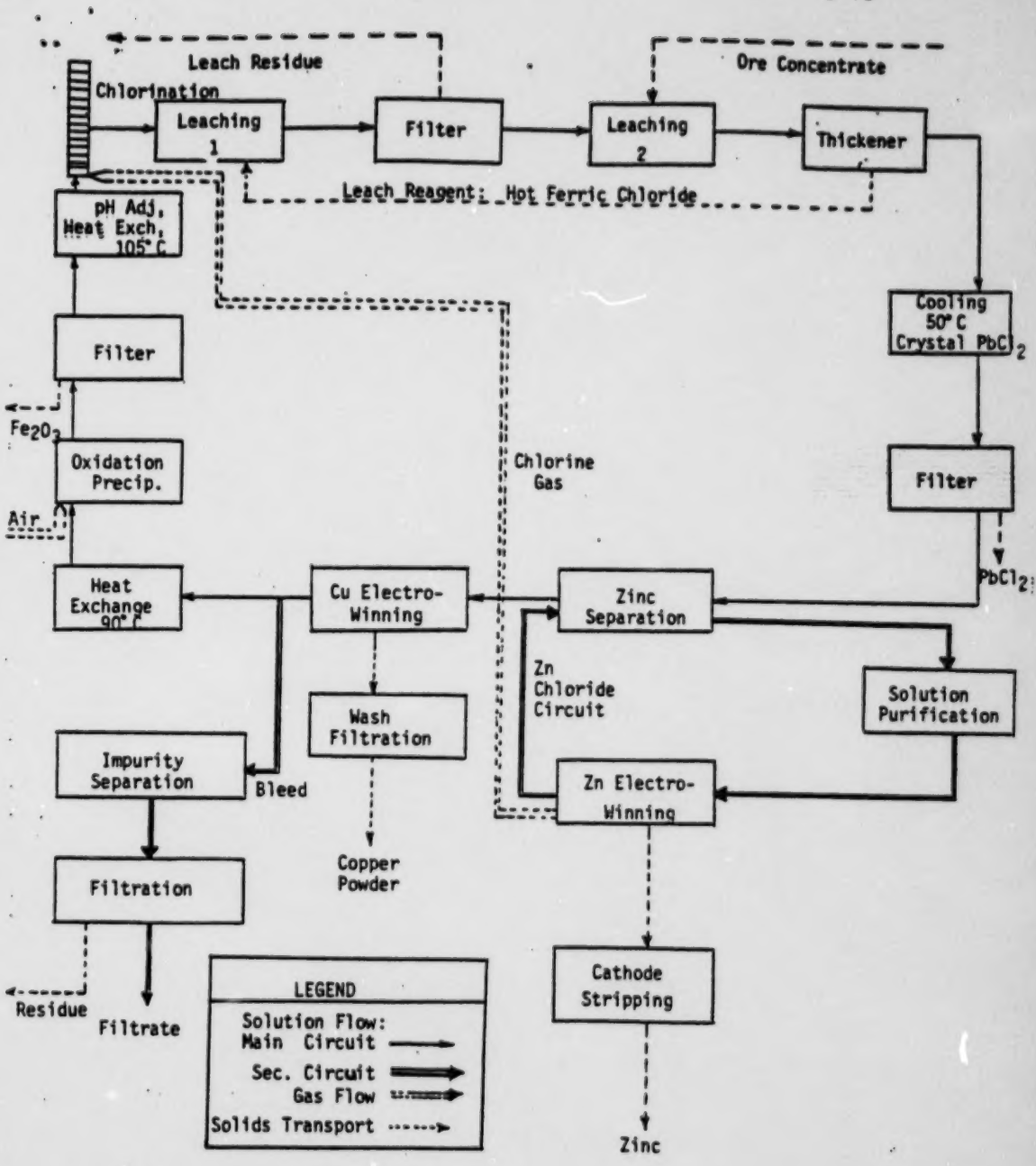


FIGURE 3.1.5A

Flowsheet of Hydrometallurgical Elkem Copper-Zinc Process

ensure that the copper is in its highest valency state, leaching of the concentrate is performed using only ferric chloride at a lower chloride concentration. Silver is recovered in a separate section.

The modified process has an advantage of producing high quality copper from dirty concentrates. The process is complicated and needs further demonstrations to enable detailed technical and economic analysis.

iii) Minemet Recherche Process (9)

This is a cupric chloride leach process (see figure 3.1.6) in which the iron leached from the feed concentrate is completely removed by precipitation of goethite by air oxidation of ferrous chloride in the presence of a mixture of CuCl and $CuCl_2$. The process has two unusual features; a divided counter-current leach and a solvent extraction circuit for converting from a chloride to sulphate circuit.

The increasing acidity resulting from goethite precipitation permits the oxidation of cuprous chloride to cupric chloride which in turn stabilizes the pH to permit continuous precipitation of geothite. The regenerated solution is returned to the second stage leach. One half of the leach liquor is sent to the solvent extraction section where half the contained copper is recovered. The remainder in the raffinate is returned to the first stage leach as cupric ions. Copper is stripped from the solvent (LIX 65 N) by spent electrolyte from a conventional copper sulphate electrowinning circuit and is recovered in that circuit as high purity cathode.

The other half of the combined leach liquors goes to reagent regeneration section. The final liquor produced in this section has the same composition as that obtained in the solvent extraction stage. The liquor is returned to second stage leach.

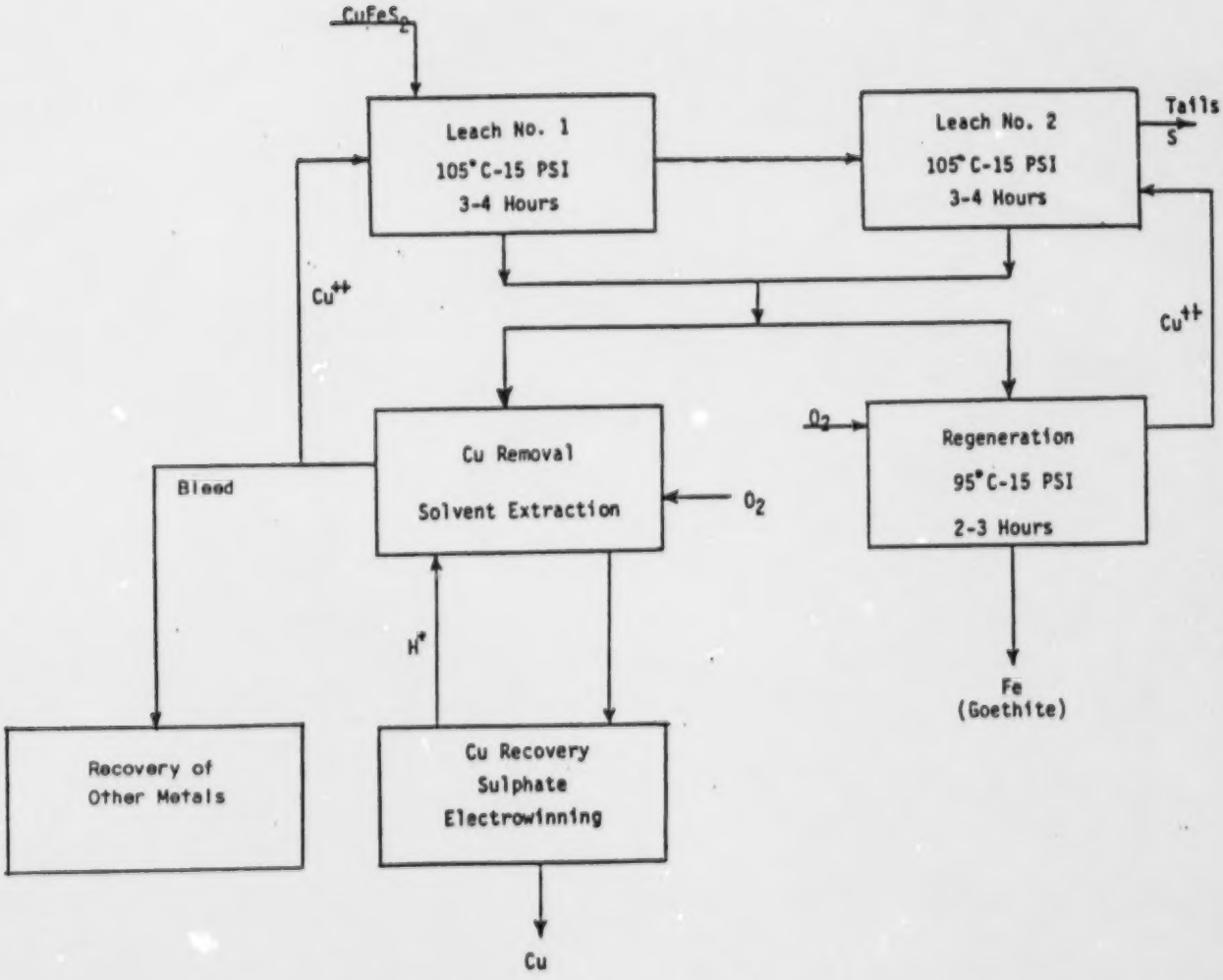


FIGURE 3.1.6

Minemet Recherche Process

Rev. 2

Zinc is extracted with DEPHA followed by solution purification and electrowinning. The process can treat both pure and dirty concentrates. Silver contamination is prevented. The economics are not known.

iv) U.S.B.M. Lead Process (11)

The process (see figure 3.1.8) involves ferric chloride leaching of galena to produce lead chlorides and fused salt electrolysis of the lead chloride to produce lead and chlorine. The chlorine produced by electrolysis is used directly to regenerate ferric chloride in the leach solution. The leaching is conducted at approximately 95°C using ferric chloride and sodium chloride.

The process is developed to treat galena in an environmentally acceptable fashion. Energy consumption at 19.8 MBTU/T is higher than the approximately 14.8 MBTU/T for the comparable KIVCET Process (3,12). The use of a chloride (eg KCl, LiCl) other than NaCl may reduce power consumption.

3.1.6 - Miscellaneous

i) U.O.P. Process for GALENA (15)

The process includes selective dry chlorination, brine leaching and fused salt electrolysis.

The disadvantages include the need for thermal activation, pretreatment and the relatively lower reactivity to zinc and copper.

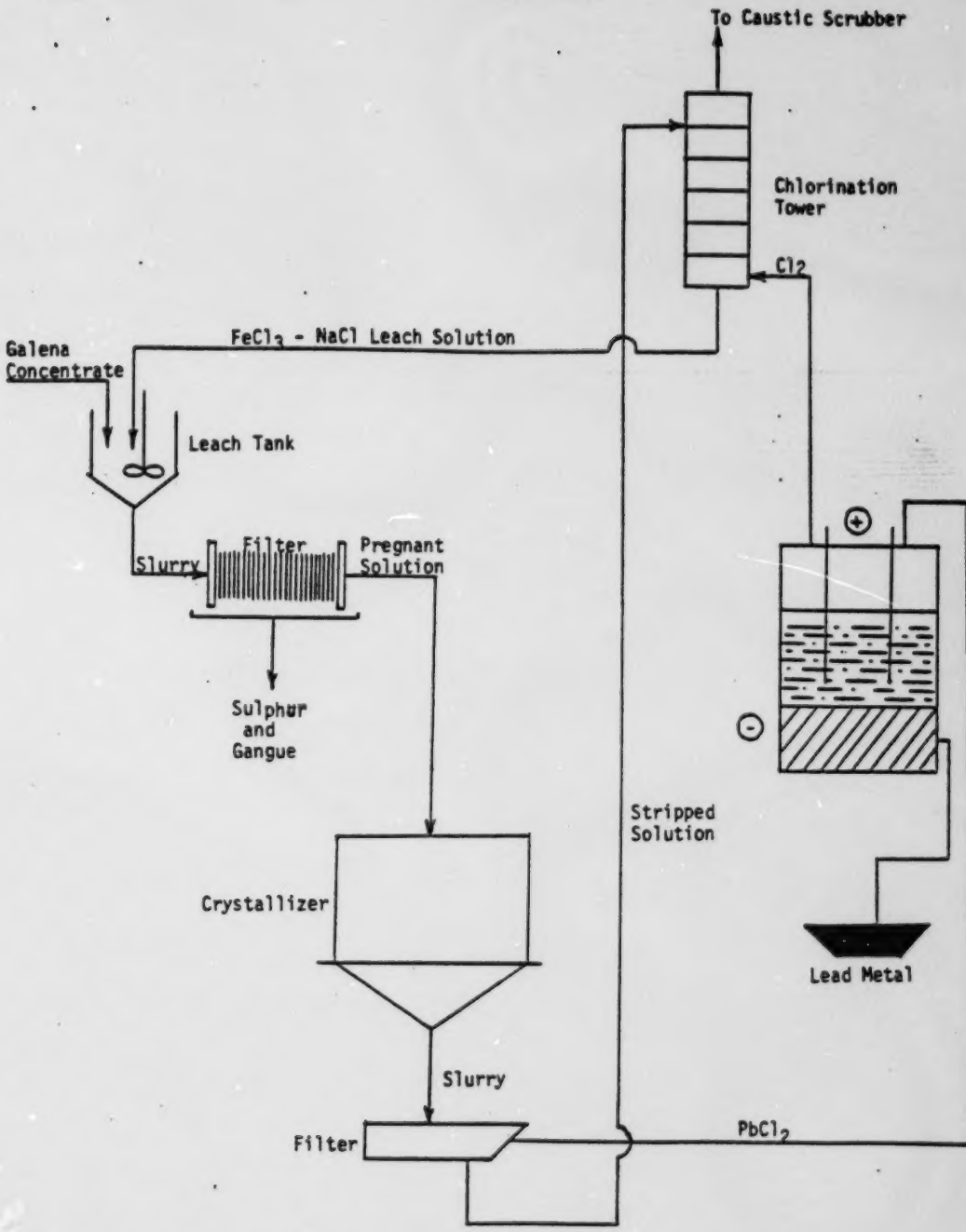


FIGURE 3.1.8

Flow Diagram of U.S.B.M. Ferric Chloride Leach/Fused-Salt Electrolysis Lead Process

ii) U.S.B.M. Ferrous Chloride - Oxygen Leaching (16)

This process (see figure 3.1.10) differed from the U.S.B.M. ferric chloride leaching process in that it proposes to convert silver and base metals to chlorides and reject iron as Fe_2O_3 and sulphur as elemental sulphur in a single step. The leach system operates without excess chlorine in order to obtain Cu-Ag separation during leaching.

iii) Zincex Process - Technicas Reunidas (11)

The process (see figure 3.1.11, 3.1.11A)is developed to treat solutions obtained from chloride roasting of pyrite cinders. The process includes two solvent extraction circuits attached to a zinc electrowinning facility. The first sx circuit uses an anionic extractant, a secondary amine, to produce pure zinc chloride solution and the second circuit includes extraction of zinc using an organo phosphoric acid. Stripping using typical tank house sulphuric acid enables normal zinc electrowinning to be used on a commercial scale. The problems associated with reagent regeneration and the recovery of other metals are not addressed.

iv) Zinchlor Process - Technicas Reunidas (17)

The process (see figure 3.1.12) includes the separation of zinc from chloride solution by a selective extractant (di-n-pentyl pentaphosphonate) in kerosene. The raffinate is recycled to leach. Zinc chloride from the loaded organic phase is stripped with hot water and is fed to Zinchlor cell after the addition of sodium chloride. The electrolysis is carried out in a diaphragm cell where the anolyte and catholyte are separated using a selective Cation Exchanger membrane. Ultra pure zinc is electrowon on a titanium cathode. Sodium ions, balancing the rate of zinc depletion, are transferred from the anolyte. Spent electrolyte from the

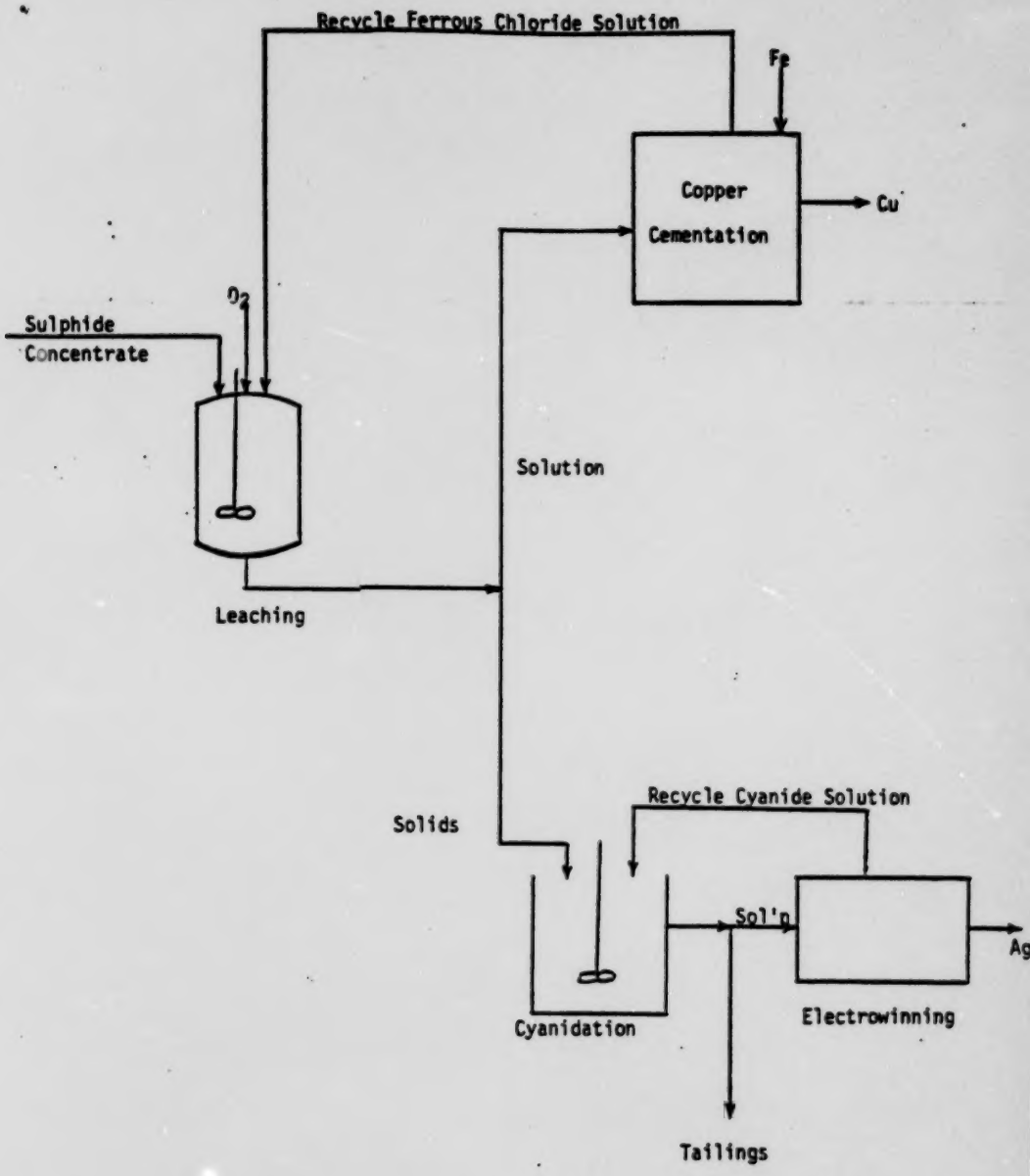


FIGURE 3.1.10

Flow Diagram for U.S.B.M. Ferrous Chloride - Oxygen Leaching of Silver-Copper Concentrate

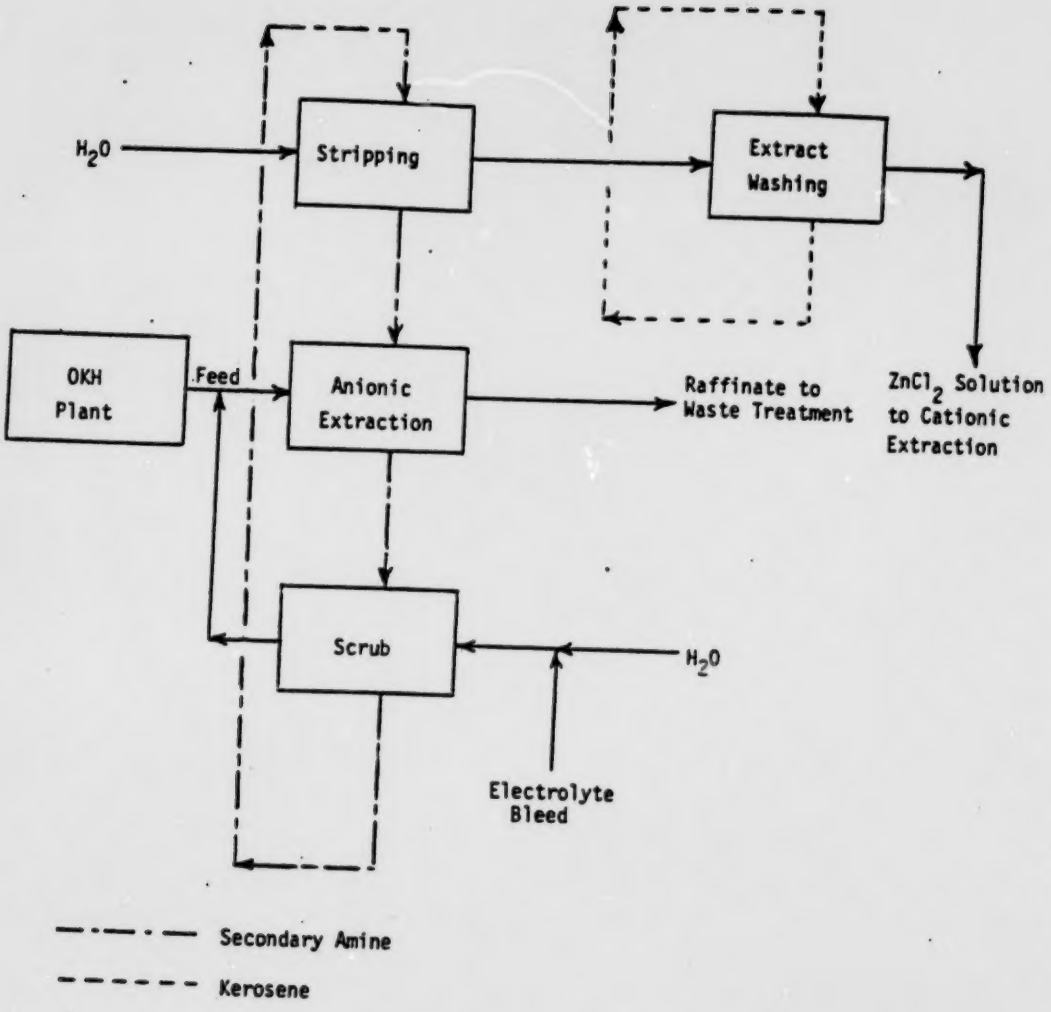


FIGURE 3.1.11

Zincex Anionic Cycle Flow Sheet

Rev. 1

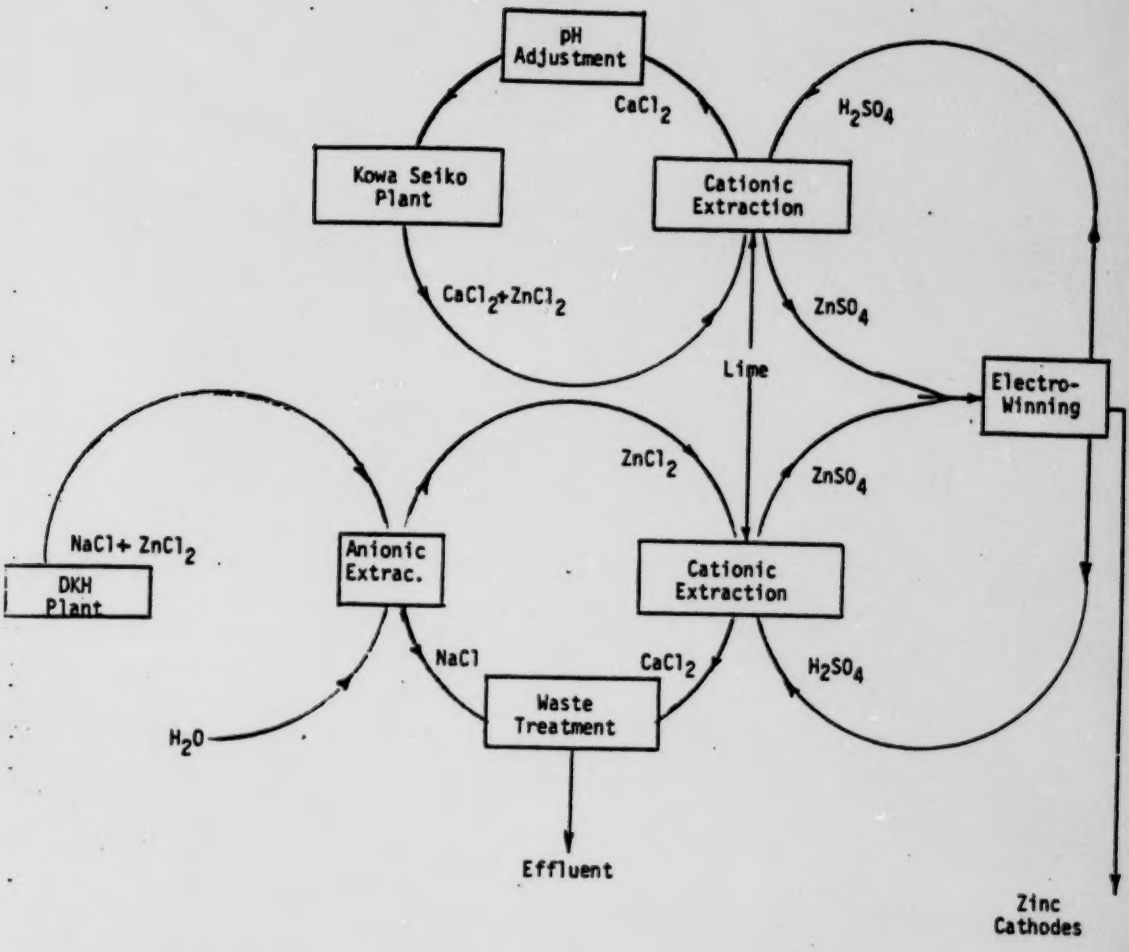


FIGURE 3.1.11A

Zincex Process
(Quimigal Plant)

Rev. 1

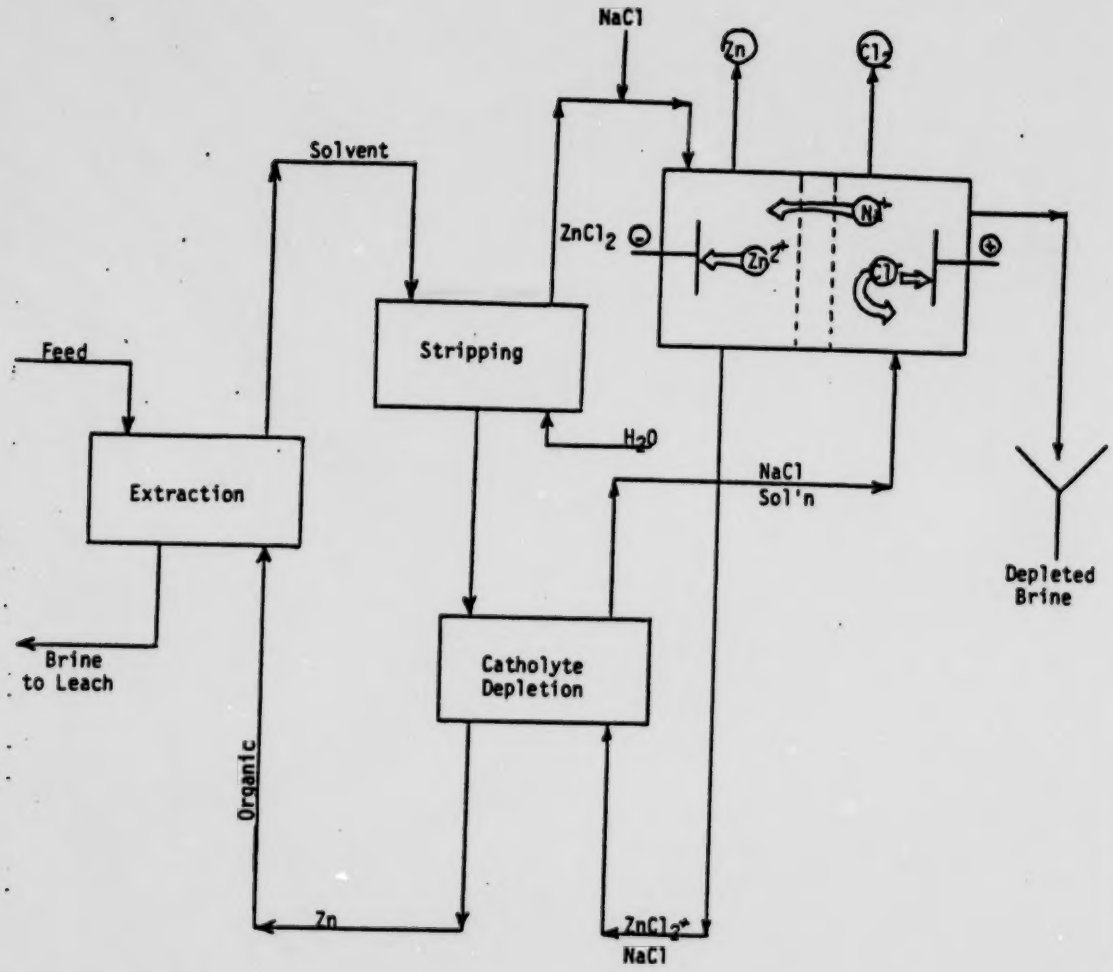


FIGURE 3.1.12

Simplified Zinchlor Process
Flow Circuits
Rev. 1

catholyte compartment, containing a low zinc concentration, is continuously transferred to solvent extraction in order to remove zinc and then returned to the anodic compartment. Pure chlorine is evolved in the anodic cell equivalent to the amount of the zinc electrowon. Depleted brine removed from anode compartment equals the amount make-up of NaCl added prior to electrowinning.

The process is proven on a pilot scale, but further details are required concerning the effect of the degradation of solvents, cathode quality, and reagent stability and availability. Due to the lower cell voltages, operating cost can be reduced. The performance of this process in an integrated circuit treating other elements is not known.

v) Cupric Chloride Leaching of Complex Turkish Cu/Zn/Pb Ores University of Birmingham (18)

Bench scale cupric chloride leach testwork to treat Cayeli Turkish ore using $CuCl_2$ as a leachant under boiling conditions revealed that grinding to less than 5 microns is necessary. Precipitation of copper can be due to the reduction of $CuCl_2 \rightarrow CuCl \rightarrow Cu^{\circ}$. A flowsheet to recover component elements has not been developed.

vi) Ferric Chloride Leaching - Indian Chalcopyrite Concentrates (19,20)

Leaching of the chalcopyrite concentrates using acidic ferric chloride at approximately 105°C was investigated. A leaching efficiency of 99 percent Cu extraction was obtained. A detailed flowsheet to recover component metals including silver is not shown.

vii) Chloride Leaching of Cobalt-Nickel Sulphide Concentrates (21)

Amax in bench and pilot scale developed a chloride process to recover cobalt and nickel from pretreated concentrates. Cu, Zn, and Fe are removed from leach solutions by TBP, and Co is separated by tri-iso-octylamine (TIOA). Cobalt is crystallized as the chloride. Ni from the raffinate is pyrohydrolyzed to the oxide.

viii) Leaching with Chlorine Solution

Sherman and Strickland (72) and Jackson and Strickland (22) made a kinetic study of dissolution of base metal sulphides in acid chlorine solutions (up to 2 g/L Cl_2) at 10 - 50 micron. They found that pyrite, pyrrhotite and arsenopyrite produced sulphate while chalcocite and covelite produced elemental sulphur. In view of the leachability of pyrite, the method does not appear to be promising for pyritic lead-zinc concentrates.

ix) U.S.B.M. Copper Process (13)

U.S.B.M.'s flowsheet (see figure 3.1.9), developed by Haver and Wong, consists of leaching with concentrated acid-ferric chloride solutions (160 g/L Fe^{3+}) at approximately 106°C, with a $FeCl_3/CuFeS_2$ ratio of 2.7: 1, and a contact time of 2 hours. Copper is electrowon in a diaphragm cell, leachant regeneration is accomplished by direct oxidation with air or oxygen and elemental sulphur recovery is carried out by treating the residue with $(NH_4)_2S$. Tests showed that As, Sb, Pb, Zn, Bi, Ni, and Ag also dissolved to varying degrees. Gold was not extracted. Ferric chloride regeneration was achieved in an oxygen-sparged, turbine - agitated atmospheric reactor at the rate of 37 gFe^{2+}/L per hour. The process was one of the first carried out using ferric chloride as a lixiviant.

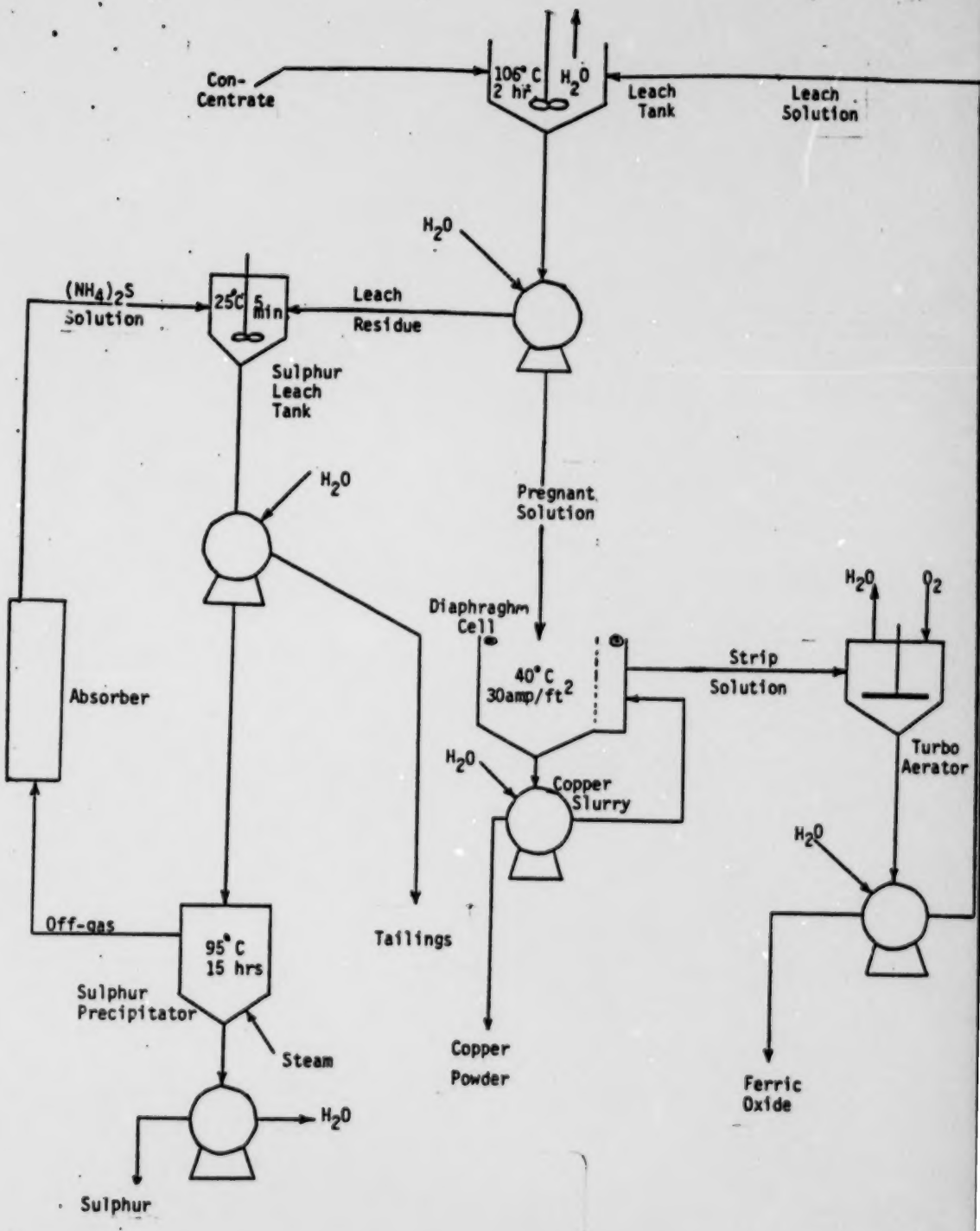


FIGURE 3.1.9

U.S.B.M. $FeCl_3$
Copper Circuit
Rev. 1

x) C.S.I.R.O. Process (14)

The process includes activation of chalcopyrite concentrate with sulphur at 400°C, followed by a 3 stage counter-current cupric chloride leach at 106°C in 1 hour. Copper from the chloride medium is recovered in a diaphragm cell. Current efficiency is only about 70 percent. The application to treat dirty complex mixed sulphides is not discussed.

3.1.7 - Patent Literature

A patent literature search of the more recent processes for the chloride extraction of copper, lead or zinc has revealed about a dozen U.S. patents and a number of Canadian and foreign patents.

Again, as with both the general process literature and the U.S.B.M. reports, most patented processes failed to deal with the issue of treatment of complex sulphides or of the disposal of waste streams bearing toxic agents. Some streams were recycled but the inevitable build-up of minor elements and the subsequent treatment of the required bleed stream was not addressed.

Some patents included pre-roasting processes (i.e. U.S. patent 4,168,969). Most chloride processes precipitate unwanted iron by solution oxyhydrolysis which also regenerates desired ferric ion for recycle to leach.

In order of ease of dissolution in an oxidizing ferric and/or cupric chloride solution medium, lead is easiest, followed by zinc, then copper. Copper is essentially present as chalcopyrite ($CuFeS_2$) which is difficult to attack, and in so doing, associated iron is also dissolved. Fortunately, pyrite is the least attacked of the associated sulphide minerals.

Canadian patent No. 1012089 shows how lead can be electrowon from a chloride solution using porous diaphragm cells to prevent ferric (regenerated on the anode) from migrating back to the cathode, thus decreasing current efficiency. Most papers that were studied examine lead recovery by molten salt electrolysis of lead chloride crystals. An aqueous system, as described in this patent, may be advantageous particularly if the molten salt route is considered too exotic for commercial exploitation.

3.1.8 - Summary

(a) Based upon the above review, it appears that the application of chloride technology offers a promising alternative to pyrometallurgy in treating complex sulphide concentrates. The technology is especially attractive on a small scale for treating extremely complex sulphide ores and where environmental regulations limit sulphur dioxide emissions.

Competition against pyrometallurgical methods on a larger scale, in treating less complex sulphide concentrates, is doubtful. The method may find an application in this case in upgrading concentrates prior to pyrometallurgical processing, or in treating very low quality concentrates, particularly with high quantities of impurities and precious metals.

The use of chloride methods to pretreat concentrates on a commercial scale is currently practised (e.g. dirty chalcopyrite (3), molybdenite concentrates).

The availability of polypropylene, kynar- and teflon-lined tanks and pipes, titanium and titanium-lined autoclaves, high temperature (80°C) rubber linings, composite plastic pumps and valves and epoxy coatings for concrete and steel tanks has assisted in materials selection for chloride metallurgy.

(b) The processes available in the literature, even under laboratory conditions, do not cover the handling of a wide variety of concentrates. Even when complex sulphides are handled, the effect of high concentrations of copper is not known.

(c) Capital and operating costs including energy and raw materials production, transportation, maintenance, process control and waste disposal, particularly of effluents containing high chloride levels, will require detailed evaluation.

3.2 - CANMET Literature Review

As part of the project, a number of internal reports prepared by CANMET staff were reviewed.

The following review is divided into sections on leaching, solution treatment, product recovery, reagent regeneration and miscellaneous.

3.2.1 - Leaching

Several reports on the leaching of sulphides, including bulk concentrates obtained from complex sulphides, with ferric chloride/sulphate were received (25, 26). The results of these reports are as follows:

In ferric ion media, the leaching rate of chalcopyrite is proportional to the surface area. The kinetics are linear for both natural and synthetic chalcopyrite. The difference in activation energies can be correlated to the difference in concentrations of $FeCl_3$, HCl, or NaCl or the presence of copper sulphide impurities. The rate is insensitive to concentrations of HCl, $FeCl_2$ or inert chlorides.

The presence of hydrochloric acid inhibits iron hydrolysis. Small amounts of $CuCl_2$ accelerate the leaching rate. Small amounts of sulphate in ferric chloride leaching reduces the chalcopyrite leaching rate. With larger amounts of sulphate, the leaching system can be compared to the ferric sulphate system. The information and relevance of lead jarosite is studied. The leaching rate in chloride media is more rapid than in sulphate media.

Percolation leach studies on pyrite Zn-Pb-Cu ores showed that the leaching rate is controlled by the rate of delivery of ferric chloride as the oxidant. Pb extraction increases with chloride ion

concentrations. Zinc extraction is relatively insensitive to HCl, $FeCl_2$ or $CuCl_2$ concentration. $PbCl_2$ precipitation causes problems during leaching. The principal variables in leaching include temperature, leachant flowrate and ferric ion concentration.

The ferric chloride leaching of sphalerite indicated that the leaching is chemically controlled. The leaching rate increases in the presence of dissolved $CuCl_2$. $PbCl_2$ does not appear to affect the reactions even though accumulations of $FeCl_2$ retard the leaching reaction. Dissolved $ZnCl_2$ impedes the reaction slightly. Low HCl concentration (<1M) had little effect on the leach rate at 50-85°C. The sensitivity of the leach increased at higher HCl concentrations.

The initial ferric chloride leaching results on complex metal sulphides indicate the order of dissolution to be galena>sphalerite> chalcopyrite>tetrahedrite. Galena is readily soluble in ferric chloride. Sphalerite and chalcopyrite leaching rates increase with increasing ferric chloride concentration.

The optimum liquid-solid ratio was 2L of 2.0M $FeCl_3$/325g concentrate. The presence of HCl was reported to suppress the hydrolysis of iron. A conceptual flowsheet designed by Dutrizac and Craigen included a first stage leach at approximately 100°C followed by a hot L/S separation. The leach liquor is contacted with excess fresh concentrate to react with residual ferric ion, as the presence of excess Fe^{3+} causes excessive Pb powder consumption during Ag cementation.

If the objective is to approach 0 g/L Fe^{3+} in the leach liquor, the lack of driving force can make chalcopyrite and tetrahedrite extractions less efficient. This is of particular concern when more than 90 percent Ag extraction is required and substantial amounts of Ag - Cu are locked in concentrate grains. Initial rapid leaching of Pb and some extraction of pyrite (even at less than 15 percent) may result in coating of the leach matrix with $PbSO_4$ or S° and can lower the efficiency of leaching of Ag and Cu.

Several leach tests were conducted by Lucas and Shimano (48). These included counter current leaching, chlorine assisted leaching and direct two stage ferric chloride leaching. Based upon the test results direct two-stage leaching is considered as a preferred route for the for the following reasons

- high Pb, Cu, Ag, Zn extractions
- low SO_4 generation in leach liquor
- low residual Fe^{3+} in leach liquor.

The recycled leachant is oxidized by chlorine prior to its addition to the leaching stage instead of being oxidized in-situ. The preferred leach conditions from these tests are

Stage 1

Charge = 202.5g of bulk conc/1.25 L of solution
Leachant = 3.5M NaCl, 0.43N HCl and 50g Fe^{3+}/L
Retention Time = 3 h
Temperature = 108°C

Stage 2

Leachant: = 50 g Fe^{3+}/L, 3.5 M NaCl and 0.43 N HCl
Retention Time = 4.5 h
Temperature = 108°C

The use of $CaCl_2$ or $BaCl_2$ to control sulphate concentration was suggested. NaCl is used to maintain a higher temperature of approximately 108°C and improve the solubilities of Pb, Ag and Cu. After optimizing, a conceptual flowsheet using direct two stage leaching was recommended, Figure 3.3.1.

3.2.2. - Solution Treatment (28-32)

The extraction and separation of copper and zinc from brine leach solution from low temperature chlorination of lead-zinc sulphide concentrates has been reported. The chosen extractants were Acorga P 5300 for copper and TBP for zinc.

Solvent extraction studies on the ferric chloride leach liquor obtained from the roast leach process showed that the liquor was too acidic to extract and separate Cu from Zn. An approach was made to extract Cu and Zn in one stage using TBP. The strip liquor from this extract will be used to separate copper from zinc. The presence of Fe^{3+} interferes with the method.

3.2.3 - Product Recovery

(a) Electrowinning (33-39)

Detailed electrowinning tests to recover Zn and Cu from chloride solutions have been conducted. Smooth, compact, dentrite free zinc deposits were obtained from zinc chloride solution containing 15 g/L, 0.12 N HCl and 15 ppm tetrabutylammonuim chloride (TBA Cl), at approximately 30 Amps/ft^2 and 30°C, using a diaphragm cell with catholyte mixing by air agitation and Dynel cloth as the diaphragm material. A current efficiency greater than 96 percent was obtained.

A relative ranking of impurities in terms of their deleterious effect is

$$Sb > Ge > Cu > Pb > Cd > Ca > Ni > Fe > Co$$

The electrolytes used were obtained from synthetic feed liquors obtained after solvent extraction separation.

Similar smooth, compact, dentrite free copper deposits were obtained from aqueous chloride solutions using diaphragm cells.

Jaguar C-13 at 0.25 - 0.5 ppm was used as an additive.

(b) Cementation

Tests carried out to study the effect of $ZnCl_2$ concentration on Cu, Cd or Pb cementation from synthetic solution showed there was negligible effect. The rates of Cu, Cd and Pb cementation were

similar and increased with Zn dust concentrations up to 8 g/L Zn dust. The quality of electrolyte obtained after Zn dust cementation was acceptable for Zn electrowinning.

Removal of Co, a deleterious impurity in Zn electrowinning, by Zn dust cementation was found to be temperature and time dependent. Increasing temperature from 35 - 70°C and time from 20 - 60 minutes favoured Zn cementation. NaCl at 3 M favoured Co cementation but the rate decreased at 30 g/L $ZnCl_2$.

Cementation of silver with copper required a nitrogen atmosphere. The presence of Fe^{3+} ions required pretreatment with granular Pb for reduction prior to silver cementation. Temperature and agitation did not effect Ag cementation on copper or lead. The presence of $CuCl_2$ had a favorable effect on silver cementation.

3.2.4 - Reagent Regeneration

The only report made available on this subject showed that the presence of NaCl (0.27 M) had a favorable catalytic effect on Fe^{2+} oxidation by molecular oxygen. The oxidation rate increased with temperature up to 90°C, above which the rate decreased.

3.2.5 - Miscellaneous

Solubility values for $PbCl_2$, AgCl in ferric chloride solutions were reported.

3.2.6 - Conclusion

Ferric chloride leaching conditions sufficient to specify design criteria for pilot plant design have been obtained for New Brunswick bulk concentrates.

Lead recovery from lead chloride has not been assessed and published data will have to be used.

Solution treatment data to remove Zn and copper are not available and published data will have to be supplemented. Recovery of Zn and copper from synthetic chloride solutions by electrowinning may allow the determination of design criteria required for the pilot plant. Effects of impurity levels during electrowinning will have to be presumed.

Silver cementation data for real solutions obtained from direct two stage leaching are not available and will have to be assumed.

Details on effluent treatment, reagent regeneration and materials of construction for pilot scale work will be supplemented from published data.

3.3 - Flowsheet Development

A characteristic of hot ferric chloride leaching is its ability to bring many metals into solution and the burden of separating the metals falls to the solution treatment portion of the flowsheet. Starting with the flowsheet outlined by CANMET(27), various stages in the evolution of the flowsheet are shown in Figures 3.3.1 through 5. The metals are separated from solution in the order: silver-lead- zinc-copper.

The objective of high metal recovery is substantially met. The final flow diagram is shown on Figure 3.3.6. References appear in the appendix.

3.3.1 - Leaching

The leach conditions of Lucas and Shimano (27,40) were accepted. This two-stage countercurrent leach, using a rather low ferric chloride concentration, provides low sulphur oxidation as well as high levels of metal extraction. Most of the sulphide sulphur, excluding that in unreacted pyrite, is converted to elemental sulphur. Sodium chloride is employed to provide the chloride concentration necessary for complexing lead and silver, and to allow a high operating temperature.

By limiting the amount of ferric chloride to the 1st leach stage to less than the reducing power of the concentrate, a leach solution low in ferric iron can be produced; this is important in the subsequent processing of the solution. Metals extraction for the concentrate is completed by further contact with ferric chloride in the 2nd leach stage. It is claimed that low ferrous ion in the lixiviants will result will result in low ferric ion in the product liquor.

3.3.2 - Silver Recovery

Because of its limited solubility at ambient temperature, lead chloride must be removed early in the recovery circuit. However,

to avoid contamination of the lead, silver must be recovered prior to lead chloride crystallization (41). Thus silver removal from hot leach solution is the first process of the recovery circuit.

Silver cementation with lead, zinc, or iron would result in a mixed silver-copper-lead product (42,43). Discussions with potential refiners revealed that a copper-lead mixture is of little interest; however a silver-copper product would be of value. The silver is therefore removed by cementation onto copper, with copper passing into solution. A discharge solution containing about 3 mg Ag/L is expected. The consequential alterations to the CANMET flowsheet are shown in Figure 3.3.2.

The copper powder will also reduce any ferric iron that is present to ferrous. Ferric is held to a low concentration, assumed at 0.5 g/L, by control of oxidation conditions in the leach. Cupric ion in the leach liquor will also be reduced and consume copper powder. Reference (44) suggests a cupric: cuprous ratio of 50:50 can be obtained in the leach; a 60:40 ratio is taken for the flowsheet calculations. The copper so consumed exceeds that contained in the concentrate but the metal from both sources is recovered later in the circuit.

3.3.3 - Lead Recovery

Lead chloride is recovered by cooling the solution to 15°C, thereby exceeding its solubility. The crystallized lead chloride is filtered from the solution, washed, and dried. Lead removal is about 50 percent complete, with 12 g/l Pb remaining in solution (45). Most of this lead is recycled to leach and is not lost, but an appreciable amount does leave the circuit in the bleed solution. Cooling below 15°C would presumably decrease the solubility of lead chloride but data are not available on the possible contamination with other metals that might result. At 15°C, except for sulphate,

the purity of the crystals was satisfactory (27). To reduce the sulphate to 5 g/L, a level considered to be satisfactory (45), calcium chloride addition to the leach has been included (41).

Electrowinning of lead from aqueous chloride solution is considered insufficiently developed to permit its inclusion (46). Although molten salt electrolysis of lead chloride has not been practised industrially, considerable test work information is available (41,47,48,49,50).

3.3.4 - Zinc Recovery

Since the zinc content constitutes about 60 percent of the metal value in the concentrate, its recovery as a finished product, e.g. cathode, was considered necessary. Three solvent extraction processes have been suggested to upgrade zinc chloride in preparation for zinc electrowinning.

The Zinchlor process employs di-n-pentyl pentaphosphonate (DPPP) as a very selective solvent (17,51). Zinc is recovered by electrowinning from zinc chloride solution simultaneously obtaining chlorine for recycle. The cell includes a cationic selective membrane and claims a lower cell power requirement. The addition of sodium chloride is, however, required, and the availability, properties, and effect of DPPP or its degradation products on cathode quality are not known.

The Zincex process, which is commercially operated, involves a first cycle extraction using a secondary amine, followed by extraction with di-ethyl hexyl phosphoric acid (D2EHPA) (52,53). Electrolysis is from a zinc sulphate solution and oxygen, rather than chlorine, is released. Since chlorine is needed to convert ferrous chloride to ferric chloride for recycle to leach, this system was not chosen. There is also a danger of contaminating the D2EHPA with secondary amine.

Tributyl phosphate (TBP)-kerosene, as used by Elkem a/s, was chosen for the flowsheet (8,31). Elkem extract from a solution containing a similar concentration of lead chloride and a rather higher concentration of copper and, with the aid of a scrubbing circuit and the addition of zinc oxide and zinc powder in an electrolyte purification circuit, make acceptable zinc cathode. Cleanup of the feed solution by cementation onto zinc should not be required and therefore the step is not included. To avoid degrading the cathode quality with the presence of TBP, the strong electrolyte is passed through a carbon column. The zinc raffinate is also treated with carbon in order that TBP does not contaminate the extractant in the following copper circuit. The zinc recovery circuit is that shown in Figure 3.3.3.

Zinc metal can be obtained by electrowinning from molten chloride salt solution (54,55,56) or from aqueous chloride solution (35). For reasons of greater simplicity and following Elkem's lead, the aqueous system was chosen. Chlorine is produced in the electrowinning cell for recycle.

TBP provides a greater selectivity over iron in reduced solutions, therefore the SX system is protected from atmospheric oxygen with a blanket of inert gas. Since the solutions must be oxidized for copper solvent extraction, zinc recovery precedes that of copper.

3.3.5 - Copper Recovery

Copper cannot be allowed to build up in the circuit to any extent (by recovering it from the bleed stream, for example) without causing a rather extravagant demand for copper powder in silver cementation. Copper removal by cementation with lead, zinc or iron, or by precipitation with hydrogen sulphide, would result in an unwanted copper-lead mixture. Although solvent extraction appears elaborate for the rather minor amount of metal involved, it is capable of recovering the copper in a saleable form. Solvent extraction with LIX or Acorga reagents will provide a sufficiently low raffinate for recycle to leach (31).

Air is introduced into the SX mixers (9) to oxidize cuprous ion to cupric, as is necessary for the copper extractants. The hydrogen ion released from the solvent is thereby made available to reduce the pH rise during the oxidation and prevent the precipitation of ferric compounds.

In addition to not producing a finished product, cementation of the copper strip solution onto iron or precipitation with sodium sulphide are not fully applicable because of the high acid content of the strip liquor. Copper electrowinning following solvent extraction is practised commercially at a number of locations and can be considered a well-proven operation requiring little development. However, since the amount of copper is rather small, the production of copper sulphate may be attractive and this route was chosen as the more informative option for the pilot plant. Although a commercial flowsheet (57) includes an evaporation step, examination of the phase diagram of the copper sulphate pentahydrate-sulphuric acid-water system (58) would indicate cooling alone should suffice in this case.

3.3.6 - Bleed Treatment

The volume of solution bled from the circuit is based on the minimum required for solution balance; this was estimated to be 1.0 m^3/t concentrate.

The bulk of the heavy metals is removed by oxidation and neutralization with lime to pH 9.0. Treatment with sodium sulphide is expected to reduce any remaining metals to sufficiently low levels for disposal, although further treatment with starch xanthate, for example, may prove to be necessary. The solution will then carry soluble chlorides, principally those of sodium and calcium. To avoid possible releaching of the hydroxides by oxidation of sulphides, the two sludges would be disposed of separately.

The iron in the bleed stream just balances that dissolved in the leach so that iron removal through an in-circuit oxidizing-precipitation process is not required. The flowsheet to this point is shown in Figure 3.3.4.

It may prove possible to neutralize to a lower pH to precipitate and dispose of the bulk of the iron, and then recycle to leach the metals precipitating on further lime addition. However, since the amount, deportment, and effect of a number of impurity elements is not known, such variations were not pursued.

3.3.7 - Chlorination

Chlorination of ferrous chloride to ferric chloride in the lixiviant is accomplished in two steps corresponding to the two leach stages. The chlorine bearing off-gases from zinc electrowinning plus make-up chlorine are passed through the 2nd stage leach chlorinator and then to the 1st stage leach chlorinator. The iron in the leach solution to the 2nd stage is thereby fully converted to ferric. The chlorine makeup rate is controlled to similarly convert all ferrous ion in the first stage lixiviant to ferric.

The routing of the chlorine is shown in Figure 3.3.5.

3.3.8 - Leach Tails

Second stage leach discharge filter cake is washed to recover soluble metal values, the filtrate passing into the leach system. Raffinate that otherwise would add to the bleed stream flow is employed. In order that the solids may be safely disposed of, the raffinate in turn is washed from the solids using water. This filtrate is passed to the bleed treatment for neutralization, etc.

Elemental sulphur in the tails solids is potentially recoverable but the flowsheet does not attempt the operation.

3.3.9 - Water Balance

The overall plant balance is shown in Section 4.2.11. The major exit streams result from copper solvent extraction scrubbing and regeneration stages; these flows are subject to reduction depending on operating results. The scrubbing and regeneration streams, however, do not enter the main circuit and carry little metal. The most important input is the zinc scrub water which increases the bleed stream proportionately.

3.3.10 - Reagent Consumption

The calculated reagent consumptions are listed in Section 4.2.12. The most significant reagent requirement is hydrochloric acid. The chlorine consumption includes that "wasted" from lead chloride reduction; in an operating plant this chlorine would be recovered and the net consumption would be less.

3.3.11 - Future Developments

The current flowsheet includes several steps where significant improvement may be possible.

(a) Silver Cementation/Copper Recovery

The addition of copper powder to cement out the contained silver inevitably also reduces any ferric and cupric ions that are present. The copper powder consumed appears as cuprous ion. The increase in copper strength during silver cementation therefore depends in part on the copper strength entering cementation and the copper entering cementation depends in part on the recirculating load of copper back to the leach with the raffinate.

If the reduction of cupric ion during silver cementation were achieved by means other than by copper powder, then the entering copper concentration could be allowed to increase and consideration could be given to the recovery of copper from the much smaller bleed stream rather than from the main solution flow. Electrolysis is suggested for evaluation as a method of reducing the cupric ion (and ferric?) in the cementation feed liquor.

(b) Zinc Solvent Extraction

Tributyl phosphate (TBP) is not highly selective for zinc chloride; the loaded organic requires scrubbing and the strip solution requires purification. A significant volume of water is added to the system because of the scrub operation, with a corresponding increase in the volume of the circuit bleed.

Di-n-pentyl pentaphosphonate (DPPP) (17,52) is reputed to be considerably more selective than TBP and may allow the immediate entry of the strip solution to electrowinning, without solvent scrubbing. Detailed study would be required to establish DPPP as a competitive extractant.

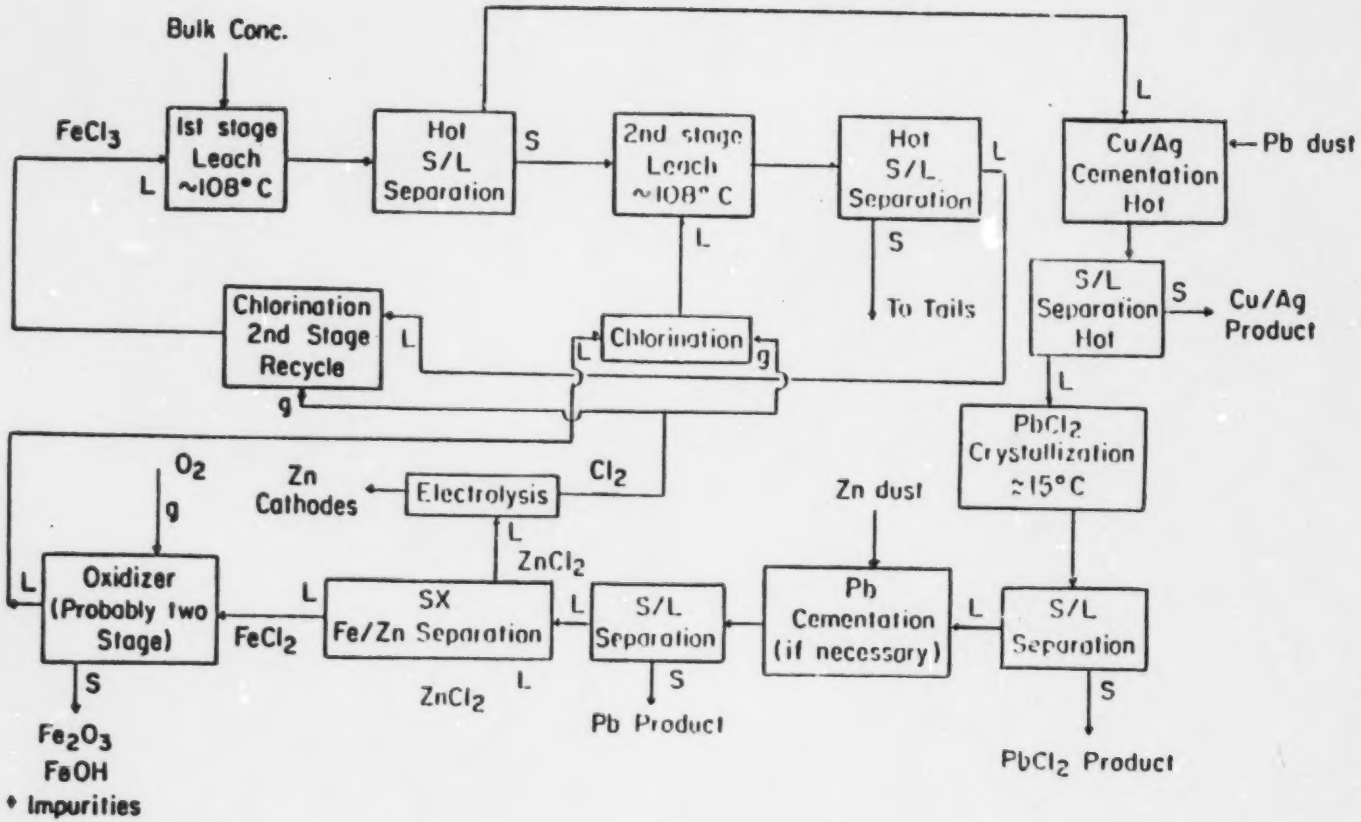
(c) Zinc Electrowinning

Zinc electrowinning from a chloride salt, rather than from sulphate, was chosen in order to produce the chlorine gas required in the leach system. This arrangement carries the penalty of enclosed anode compartments (to collect the chlorine), scrubbing of the spent electrolyte, and equipment for handling and using the collected chlorine gas.

The recovered chlorine is employed to convert ferrous chloride to ferric chloride. However, providing hydrochloric acid is present, this can also be accomplished by oxidizing the ferrous solution with air. The production of a zinc sulphate solution for electrowinning, while converting the chloride into hydrochloric acid, may be possible using a two-cycle extraction system similar to the Zincex process (52,53).

(d) Bleed Solution Treatment

The volume of the bleed solution has been set by circuit water balance considerations. The contained sodium, iron, lead, zinc, copper, and minor undefined impurities are thereby lost from the system. Whereas the impurities require purging from the circuit, the lead, zinc, and copper content represents a loss; iron is in precarious balance. By neutralizing to various pH end points, it may be possible to return much of the lead, zinc, and copper to leach for eventual recovery, to retain a controlled amount of iron, and to bleed sufficient minor metals as required for circuit purity.



PROPOSED TWO-STAGE FERRIC CHLORIDE LEACH PROCESS FLOWSHEET.

Figure 3.3.1
Proposed CANMET Flow Sheet

TWO-STAGE LEACH

Silver cementation altered

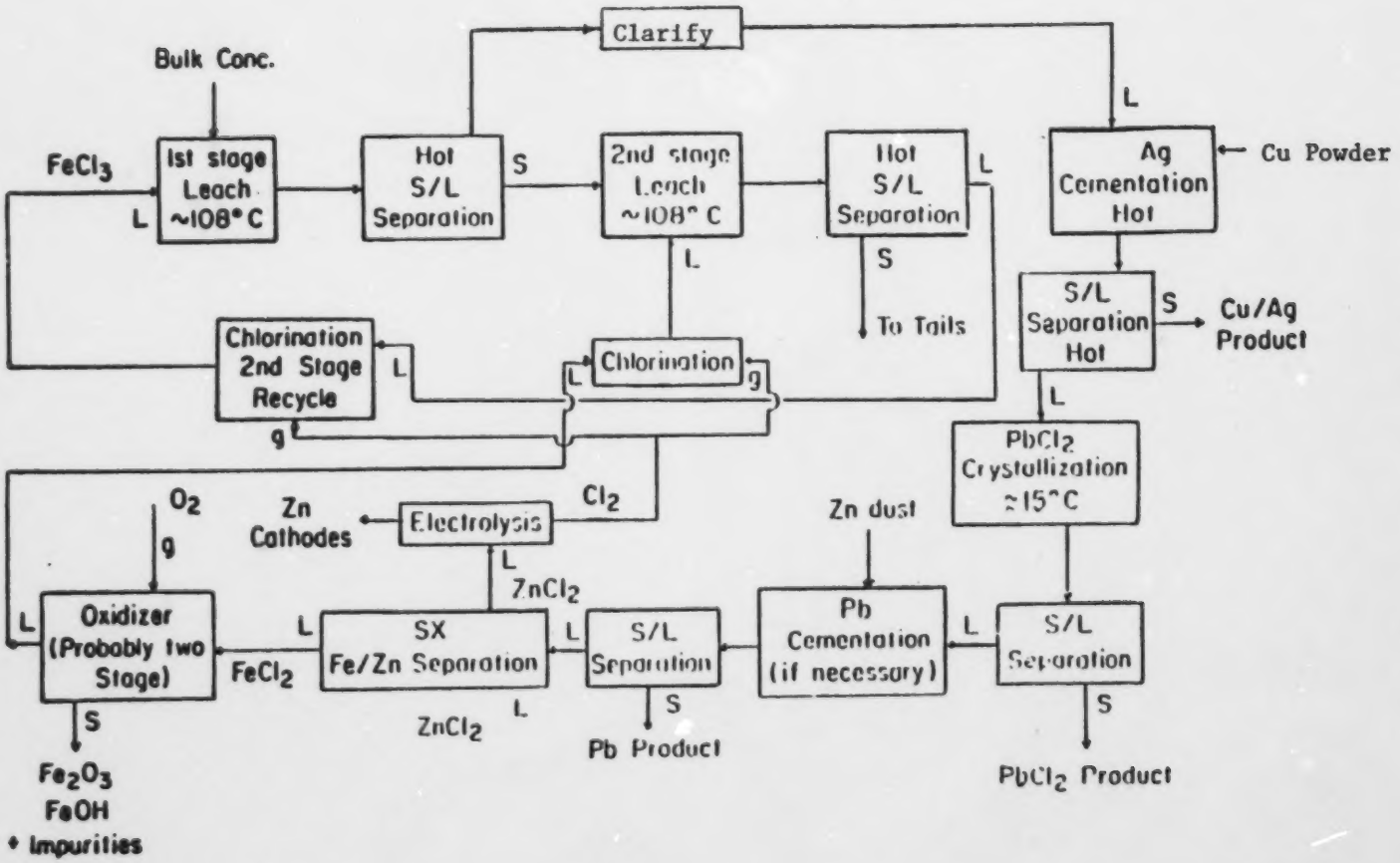


Figure 3.3.2
Silver Cementation with Copper

TWO-STAGE LEACH

Silver cementation altered
Lead cementation deleted

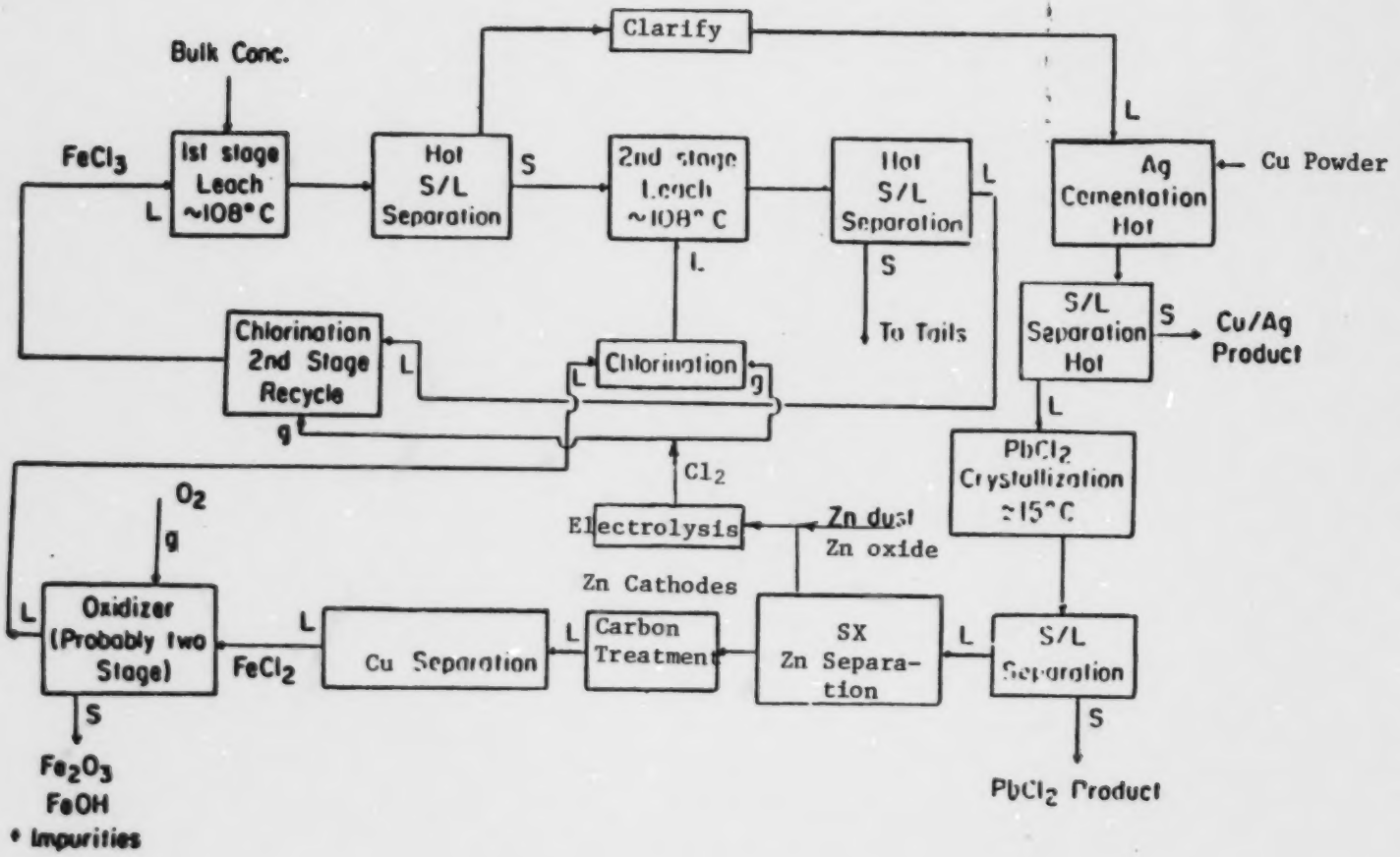


Figure 3.3.3
Lead Cementation Deleted

TWO-STAGE LEACH

Silver cementation altered
Lead cementation deleted
Copper SX added
Iron removal deleted

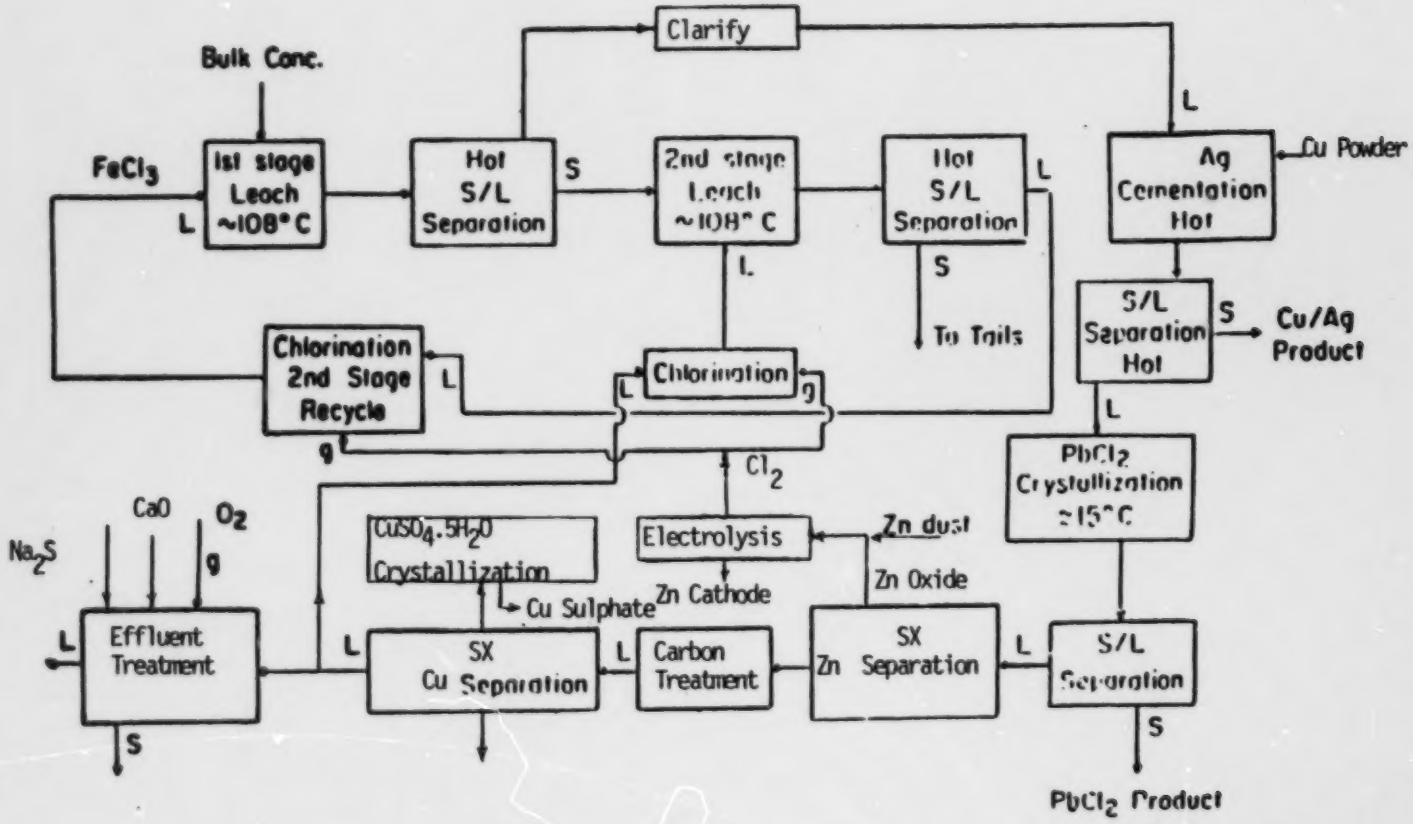


Figure 3.3.4
Copper SX and Bleed Treatment

Silver cementation altered
Lead cementation deleted
Copper SX added
Iron removal deleted
Chlorine flow altered

TWO-STAGE LEACH

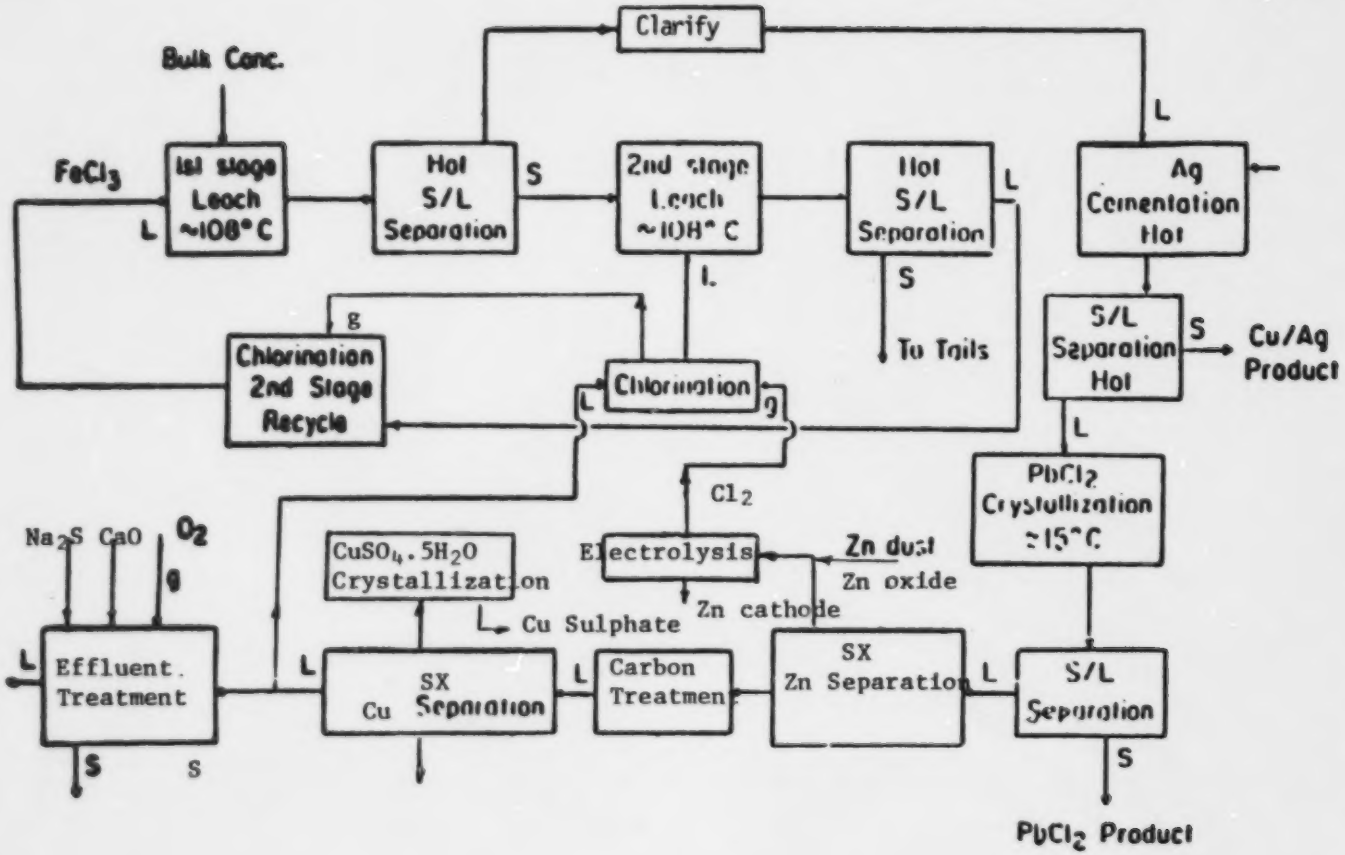


Figure 3.3.5
Two-Stage Chlorination

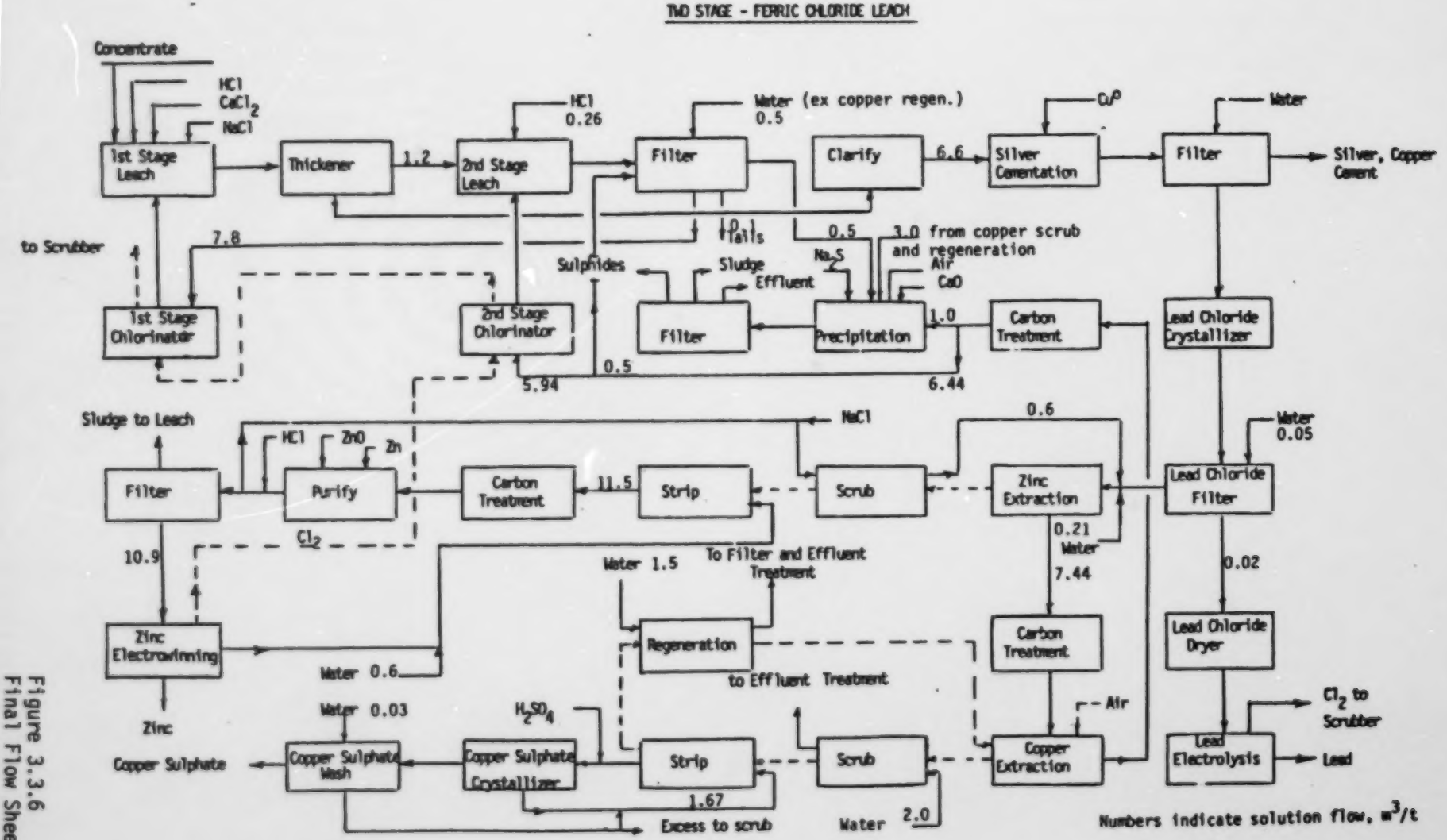


Figure 3.3.6
Final Flow Sheet Rev.1

4 - PILOT PLANT DESIGN

4 - PILOT PLANT DESIGN

4.1 - Design Basis

4.1.1 - Plant Capacity

The mini-plant will be designed to treat 10 kg/h of dry concentrate.

4.1.2 - Plant Availability

The mini-plant will be designed to operate continuously on campaigns of seven days or more on the basis of two 12 hour shifts per day.

4.1.3 - Ore Specification

The mini-plant will be designed to treat concentrate of the following specification

Analysis

Element	Percent
Pb	8.75
Zn	28.1
Cu	0.78
Ag	233 g/t
Fe	19.1
ST	35.0
S/SO4	1.51
Other	8.27
Moisture Content	7.0
Density	5 t/m3

4.1.4 - Metals Leach Extractions

The mini-plant will be designed on the basis of the following metals leach extractions.

Element	Percent
Zinc	98
Lead	99.5
Copper	93
Silver	91

4.1.5 - Raw Materials, Reagents and Services

The mini-plant will be designed on the basis of the supply of the following raw materials, reagents and services at the plant battery limits.

(a) Concentrate

Method of supply	In drums
Storage required	21 days minimum

(b) Hydrochloric Acid

Composition	32% HCl
Method of supply	In drums
Storage required	21 days consumption minimum

(c) Calcium Chloride

Form	Powder
Method of supply	In bags
Storage required	21 days consumption minimum

(d) Sodium Chloride

Form	Powder
Method of supply	In bags
Storage required	21 days consumption minimum

(e) Chlorine

Method of supply	In cylinders
Storage required	21 days consumption minimum

(f) Nitrogen

Method of supply	In cylinders
Storage required	21 days consumption minimum

(g) Copper Powder

Method of supply	In bags or drums
Storage required	21 days consumption minimum

(h) Slaked Lime

Form	Powder
Method of supply	In bags
Storage required	21 days consumption minimum

(i) Sodium Sulphide

Form	Powder
Method of supply	In bags
Storage required	21 days consumption minimum

(j) Tributylphosphate (TBP)

Method of Supply	In drums
Storage required	21 days consumption minimum

(k) Solvesso 150

Method of supply	In drums
Storage required	21 days consumption minimum

(l) Zinc Oxide

Form	Powder
Method of supply	In bags
Storage required	21 days consumption minimum

(m) Zinc Powder

Method of supply	In bags
Storage required	21 days consumption minimum

(n) Acorga P5300

Method of supply	In drums
Storage required	21 days consumption minimum

(o) Shellsol LX154

Method of supply	In drums
Storage required	21 days consumption minimum

(p) Sulphuric Acid

Composition	98% H_2SO_4
Method of supply	In drums
Storage required	21 days consumption minimum

(q) Potassium Chloride

Form	Powder
Method of supply	In bags
Storage required	21 days consumption minimum

(r) Lithium Chloride

Form	Powder
Method of supply	In bags
Storage required	21 days consumption minimum

(s) Caustic Soda

Composition	50% NaOH solution
Method of supply	In drums
Storage required	21 days consumption minimum

(t) Activated Carbon

Form	Granules
Method of supply	In bags
Storage required	21 days consumption minimum

(u) Process Water

Method of supply	In pipeline

(v) Process Air

Method of supply	In pipeline
Supply pressure required	690 kPag

(w) Electrical Power

Method of supply	Numerous power outlets
Power Supply	3 phase 550 V 60 Hz
	1 phase 110 V 60 Hz

(x) Low Pressure Steam

Method of supply	In pipeline
Supply pressure required	275 kPag

4.1.6 - Products, By-Products and Effluents

The mini-plant will be designed on the basis of the discharge of the following products, by-products and effluents at the plant battery limits.

(a) Silver/Copper Cement

Form	Filter cake containing 15% moisture
Method of discharge	In drums

(b) Zinc

Form	Electrowon cathodes

(c) Copper Sulphate Pentahydrate

Form	Crystals
Method of discharge	In drums

(d) Lead

Form	Lead bullion

(e) Tailings

Form	Filter cake
Method of discharge	In drums

(f) Bleed Effluent, Liquid Waste Steams

Method of discharge	In drums

(g) Bleed Filter Cake, Oxides

Method of discharge	In drums

(h) Bleed Filter Cake, Sulphides

Method of discharge	In drums

(i) Clean Off-gas

Method of discharge	In vent pipe outside building

(j) Spent Activated Carbon

Method of discharge	In drums

4.1.7 - Environmental Conditions

(a) Location

The mini-plant will be designed on the basis that it will be erected inside an existing building which is provided with electrical power, process water, steam, and process air supplies. It is assumed that the building will have adequate fire protection facilities.

(b) Plant Battery Limits

As defined on general arrangement drawing number C1016-M-001.

4.1.8 - Units

The mini-plant will be designed using metric units.

4.2 - Design Criteria

4.2.1 - Area 100 - Leaching

1st Stage

Time, total in 3 vessels	3 h
Temperature	108°C
Concentrate rate	10 kg/h
Solution rate	77.7 L/h
Hydrochloric acid, (32%)	
addition rate (normally zero)	2.6 L/h
control acidity	0.1 N discharge
Sodium chloride	
addition rate	1.05 kg/h
control level	3.0 N
Calcium chloride	
addition rate	0.75 kg/h
location	No. 1 vessel
control SO_4	5 g/L

Interstage thickener

Unit area (assumed)	0.7 m^2 d/t
Underflow solution	1.2 L/kg
Overflow solids (assumed)	500 ppm

2nd Stage

Time, total in 3 vessels	4.5 h
Temperature	108°C
Solution rate	59.4 L/h
Hydrochloric acid, (32%) rate	2.6 L/h

Cooling tank

Time	1.5 h
Discharge temperature	85°C

Tails washing	
Solids rate	5.65 kg/h
Solution rate	73.9L/h
Pressure filters, number	2 off
Cake moisture assumed	15%
Raffinate wash	0.5 L/kg concentrate
Water wash (from copper regeneration)	0.5 L/kg concentrate
Operating temperature	85°C
Unclarified surge tank	
Capacity	8 h consumption
Temperature	80°C minimum

Notes:

Foaming may be expected in the lead vessel of the 1st stage leach.

Leach vessels, thickener and unclarified surge tank are to be covered and insulated.

1st stage leach vessels, thickener and unclarified surge tank are to be blanketed with inert gas.

Filters are to be sized to hold 12 hours tails solids each and they are to be purged with inert gas before use.

Total acid addition to leach will be 2.6 L/h, entering 2nd stage. Substantially all the acid is expected to appear in 2nd stage discharge, to be consumed in 1st stage. Fresh acid to 1st stage will be minor, to control the acidity of leach discharge.

4.2.2 - Area 200 - Silver Recovery

Clarification Filter	
Pressure filters, number	1 off
Silver Cementation	
Temperature (minimum)	80°C
Stirred vessels	3 off
retention time, each	0.5 h
Copper powder, addition rate	129 g/h
Silver Filter	
Pressure filters, number	1 off
Filtrate, silver concentration	3.0 g/L

Notes:

Cementation tanks are to be covered, insulated, and blanketed with inert gas. The filters are to be purged with inert gas before placing in operation.

The filters are to be sized for cleaning once per week.

Copper powder addition rate is based on 0.5 g/L ferric and 0.8 g/L cupric ion (60% of total) in the feed solution.

4.2.3 - Area 200 - Lead Chloride Recovery

Crystallization

Batch operating time	12 h
Number of crystallizers	2
Solution feed rate	66.0 L/h
Feed temperature, approx.	80° C
Discharge temperature	15° C
Lead in feed	23.4 g/l
Lead in discharge	12 g/l

Filtration

Pressure type filters, number	1 off
Solids capacity	12 h
Wash water	
consumption	4 L/batch
acidity (HCl)	pH 1.5
Moisture in solids	16%

Zinc extraction feed tank

Capacity	16 h consumption
Temperature	approx 25°C

Notes:

The solution in the crystallizers and filters is to be protected from oxidation by blanketing with inert gas.

4.2.4 - Area 200 - Bleed Treatment

1st stage

Method	Batch operation using oxidation with air and neutralization with lime
Batch operating time	12 h
pH	9.0
Stoichiometric O_2 rate	2.0 g mols/h
Lime rate (90% grade, 90% reactivity)	1.4 kg/h
Weight of filter cake	1.5-3.0 kg/h

2nd stage

Method	Batch operation using sodium sulphide addition
Batch operating time	12 h
Sodium sulphide addition rate	0.1 kg/h

4.2.5 - Area 300 - Zinc Solvent Extraction

Extraction

Aqueous feed - flow		74.4 L/h
	Zn	43.9 g/L
	Pb	10.7 g/L
	Cu	2.5 g/L
	Fe	52.3 g/L
Raffinate, Zn concentration		4.0 g/L
Solvent, flow		229 L/h
	TBP	60%
	Solvesso 150	40%
	loaded organic	15 g Zn/L
	stripped organic	2 g Zn/L
Number of stages		6
Mixer O/A ratio		3/1
Mix time		1.8 minutes
Settler flow rate		4.4 $m^3/h.m^2$

Scrubbing

Solvent flow		229 L/h
Scrub solution, flow		6.0 L/h
	composition	Strong electrolyte
	Zn	43.4 g/L
	pH (HCl)	1.0
	NaCl	3 N

Number of stages, iron scrub	2
copper scrub	1
Mixer O/A ratio	3/1
Mix time	1.0 minutes
Settler flow rate	4.4 $m^3/h.m^2$
Stripping	
Solvent, flow	229 L/h
loaded organic	15 g Zn/L
barren organic	2 g Zn/L
Strip solution, flow	115 L/h
barren strip Zn	14.2 g/L
pH (HCl)	1.0
pregnant strip Zn	40 g/L
Number of stages	6
Mixer O/A ratio	2.5/1
Mix time	1.8 minutes
Settler flow rate	4.4 $m^3/h.m^2$
Solvent storage tank, capacity	400 L
Purification feed tank, capacity	800 L
Lean solution storage tank, capacity	200 L

Notes:

Extraction, iron scrub, and solvent storage tank are to be protected from oxidation with an inert gas blanket. For test purposes, copper scrub is to be provided with inert gas blanket or air sparge into the mixer.

4.2.6 - Area 300 - Zinc Strip Purification

Pregnant strip flow rate	115 L/h
Solvent removal	upflow carbon column
Iron, copper removal vessels, stirred	3
retention time, each	0.5 h
Zinc oxide	
addition rate	0.47 kg/h
addition point	first vessel
control pH	4.0
Zinc powder	
addition rate, allowance	20 g/h
addition point	second vessel
Clarification	
No. of filters	1
Cleaning cycle	1/24 hour

4.2.7 - Area 300 - Zinc Electrowinning

Feed tank	400 L
Stripped liquor tank	800 L
Strong electrolyte, flow	109 L/h
Zn	43.4 g/L
HCl addition (32%)	1.0 L/h
Control pH	1.0
Spent electrolyte Zn	15 g/L
Electrolytic cell	
Type	diaphragm
Current density	400 A/m^2
Temperature	ambient
Catholyte agitation	air
Zinc recovery rate	3.1 kg/h

4.2.8 - Area 400 - Copper Solvent Extration

Extraction	
Aqueous Feed	
Flow	74.4 L/h
Cu	2.5 g/L
Zn	4.0 g/L
Fe	52.1 g/L
pH	1.0
Raffinate Cu	0.3 g/L
Solvent	
Flow	84 L/h
Acorga P5300	10%
Kerosene	90%
Cu loaded	2.8 g/L
Cu stripped	0.8 g/L
Number of stages	2
Mix time	5 minute
Settler flow rate	4.4 $m^3/L.m^2$
Air injection to each mixer	
Total stoichiometric O_2	3.0 g mols/h
Operating pH	2.0

Scrub	
Scrub solution	
Flow	20 L/h
Composition	Water
Stages	2
Mixer O/A ratio	3/1
Mix time	1 minute
Settler flow rate	4.4 $m^3/h.m^2$
Stripping	
Strip solution (from crystallization)	
Flow	16.7 L/h
Cu	40 g/L
H_2SO_4	173 g/L
Pregnant strip	
Cu	50 g/L
H_2SO_4	158 g/L
Number of stages	2
Mix time	1 minute
Settler flow rate	4.4 $m^3/h.m^2$
Regeneration	
Regenerant	
Flow	15 L/h
Composition	Water
Number of stages	1
Mix time	1 minute
Settler flow rate	4.4 $m^3/h.m^2$
Mixer O/A ratio	3/1

4.2.9 - Area 400 - Copper Sulphate Crystallization

Pregnant strip feed to crystallization	
Rate	16.7 L/h
Cu	50 g/L
H_2SO_4	170 g/L
Barren strip from crystallization	
Cu	40 g/L
H_2SO_4	173 g/L
Acid addition to pregnant strip	0.27 kg/h
Crystallization temperature	
Start	25°C
Finish	10°C
Batch operating time	12 h
Filtration	
Pressure type	
Filter, number	1 off
Solids capacity	12 h
Crystallization rate	
Copper	0.17 kg/h
Copper sulphate pentahydrate	0.66 kg/h

4.2.10 - Area 500 - Lead Chloride Electrolysis

Lead chloride dryer	
Type	oven
Final moisture	0.1% max.
Temperature	200°C
Lead chloride crusher	
Type	roll crusher
Capacity	200 kg $PbCl_2$
Lead chloride feed	
Average feed rate	1.01 kg/h
Design feed rate	1.46 kg/h
Electrolytic cell	
Design current	300 A
Operating temperature	450°C
Electrolyte composition	
Lead chloride	40-70%w/w
Potassium chloride: lithium chloride	55:45 w/w

Notes:

Horizontal monopolar graphite anodes are to be employed.

The cell is to operate at a slight negative pressure with the off gases treated in a scrubber to remove chlorine.

4.2.11 - Water Balance

In	L/h
Concentrate at 7% moisture	0.8
Hydrochloric acid	2.6
Lead chloride wash	0.5
Zinc scrub water	6.0
Copper scrub water	20.0
Copper regeneration	15.0
Copper sulphate wash	0.3
Other	2.1
Total	47.3

Out	
Leach evaporation	0.8
Tails moisture	1.0
Lead chloride moisture	0.2
Copper sulphate moisture	0.3
Effluent sludge moisture	3.0
Effluent	42.0
Total	47.3

Note:

First leach stage lixiviant may be employed to dissolve NaCl and $CaCl_2$ prior to entering the leach.

Internal flows are indicated on Figure 3.3.6.

4.2.12 - Reagent Consumptions

		g/kg feed
Sodium Chloride		
leach	105	
Zn scrub	81	186
Calcium Chloride Dihydrate		
leach		99
Hydrochloric Acid (32%)		
leach	298	
leach chloride wash	negligible	
zinc scrub	132	430
Chlorine		
chlorination		70
Copper Powder		
silver cementation		13
Zinc Oxide		
electrolyte purification		47
Zinc Powder		
electrolyte purification		2
Sulphuric Acid (93%)		
copper sulphate recovery		30
Calcium hydroxide		
effluent		137
Sodium Sulphide		
effluent, allowance		10

4.2.13 - Disposition of Metals and Bleed Solution Analysis

(a) Disposition of Metals

	Pb	Zn	Cu	Ag	Fe
Concentrate	100.0	100.0	100.0	100.0	100.0
Powder	0.0	0.7	165.3	0.0	0.0
Oxide	0.0	13.4	0.0	0.0	0.0
Total	100.0	114.1	265.3	100.0	100.0
Trails solids	0.5	2.0	7.0	9.0	70.0
Trails solution	1.4	0.4	0.7	0.4	2.7
Bleed solution	11.8	1.4	3.9	1.1	27.3
Recovery (metal)	86.3	110.3	214.1	0.0	0.0
Recovery (cement)	0.0	0.0	39.6	89.5	0.0

(b) Bleed Solution Analysis

		Feed	Neutralized
Flow	L/H	10.0	45
Pb	g/L	10.34	-
Zn	g/L	4.0	-
Cu	g/L	0.3	-
Ag	mg/L	2.66	-
Fe^{+2}	g/L	45.62	-
Fe^{+3}	g/L	6.55	-
Na	g/L	66.67	14.8
H	g/L	0.01	pH 9.0
SO_4	g/L	4.44	-
Ca	g/L	-	12.9

Stoichiometric O_2 required	2.0 g mols/h
Final pH	9.0
Calculated $Ca(OH)_2$ (92% grade, 90% reactivity)	1.4 kg/h
Estimated weight of solids	1.5 - 3.0 kg/h
Na_2S, allowance	0.1 kg/h

4.3 - Process Description

4.3.1 - Area 100 - Leaching Circuit (PFD-001)

The concentrate is received in drums and emptied into a hopper from which it is fed to the first of the three 1st stage leaching vessels, 100-V1, by means of the tubular spiral conveyor package, 100-H1. The concentrate enters the vessel through a rotary valve and is then mixed with recycle solution from the 2nd stage leach. A mixture of calc[illegible]loride (for sulphate control) and sodium chloride (for boiling point elevation) solutions is continuously fed to the leach stage from tanks 100-V12A/B via pump 100-P7. Hydrochloric acid (32%) can be added to the 1st stage leach, if required, by means of the metering pump, 100-P6.

The 1st stage leach pulp progresses, via gravity overflow, through the three 1st stage leach vessels in series, 100-V1-V3. The temperature of the leach is maintained at 108°C by steam and electrical heating of all three vessels. Nitrogen gas is used to continuously purge the vessels and to blanket the settling cone, 100-V13 and the pregnant solution tank, 100-V8. The gas enters 100-V3 and, flowing counter-currently through the overflow connections, exits 100-V1 and is drawn to the off-gas scrubber (area 500). From 100-V3, the slurry is pumped to the settling cone, 100-V13.

The overflow from the settling cone enters the pregnant solution tank, 100-V8, where it is cooled to 85°C, and is then pumped, through clarifier, 100-F1, to the silver recovery section.

The underflow from the settling cone flows by gravity, on a periodic basis, to the first of the 2nd stage leach vessels, 100-V4. The 2nd stage leach lixiviant is stripped recycle solution from the metals recovery circuit which is chlorinated in the clorinator, 100-V10, and heated to 108°C in the integral chlorinator storage tank. The solution is pumped to 100-V4. Hydrochloric acid is added to the 2nd stage leach via metering pump 100-P6.

The flow of leach pulp in the second stage leach is very similar to that in the 1st stage leach. The temperature of the leach is maintained by electric heating. From the third vessel in the 2nd stage leach, 100-V6, the slurry passes into the 2nd stage leach cooling vessel, 100-V7, where it is cooled to 85°C. From 100-V7, the slurry is pumped through one of the recessed plate tails filters 100-F2A or B. The filters operate on a continuous basis on a 12-hour cycle. The filter cake undergoes two washes; the first wash is performed using a portion of the stripped recycle solution from the metals recovery circuit; the second wash is used for chloride removal and utilizes regenerant/strip water from the copper recovery section (area 400) - this water is then directed to the bleed final precipitator in area 200. After washing, the tails are discarded.

The filtrate from the tails filters is directed to the 1st stage leach chlorinator, 100-V11, where it is contacted with chlorine exiting the 2nd stage chlorinator, 100-V10. The chlorinated filtrate is collected in the integral chlorinator storage tank from where it is pumped to the 1st stage leach vessel, 100-V1 and as make up to the salt mixing tank, 100-V12 A or B. The gas leaving the chlorinator is drawn to the chlorine scrubber in area 500.

4.3.2 - Area 200 - Silver Recovery and Bleed Treatment (PFD-002)

(a) Silver Recovery

From the leach circuit, the pregnant solution enters the first silver cementation mixing vessel, 200-V1, where it is mixed with copper powder, entering through the rotary valve from the copper powder feeder, 200-H1. The mixture flows through the three cementation vessels in series 200-V1-V3, by gravity overflow and is then pumped through the plate-and-frame cement filter, 200-F1. Temperature is maintained at 85°C by electrically heating the cementation vessels. Nitrogen gas is used as counter-current purge, as in the 1st stage leach vessels.

The copper-silver cement filter cake is water-washed once per week and recovered for sale.

The filtrate from the cement filter enters one of the lead chloride crystallizer package units, 200-X1A or B, each having a 12-hour storage capacity and operating on a batchwise basis. The solution is cooled to 15°C and is then pumped through the recessed-plate lead chloride filter, 200-F2, where the crystals are water washed once every 12 hours and recovered for fused salt electrolysis in area 500.

The filtrate from the filter enters the zinc extraction feed tank, 200-V5, from where it is pumped to the zinc solvent extraction mixer/settlers in area 300. All vessels in the silver recovery circuit are nitrogen blanketted in order to prevent oxidation. The filters are purged with nitrogen before use.

(b) Bleed Treatment

The stripped recycle solution from area 400 enters the carbon column, 200-V6, in order to remove trace organics from the recycle solution. The majority of the recycle solution then flows to the second-stage chlorinator in area 100. A small part of the recycle solution enters the wash tank 200-V9 and is used for tails filter cake washing.

The bleed stream is taken from the recycle solution after the carbon column. The stream enters the bleed oxidizer, 200-V7, where it is oxidized using process air and calcium hydroxide on a batch basis. The slurry is pumped through the recessed-plate bleed filter, 200-F3, once per day and the filter cake is discarded. The filtrate enters the bleed final precipitator, 200-V8, where sodium sulphide is added in order to precipitate any trace metals in the filtrate. The

underflow is pumped on a batchwise basis through the bleed filter and the sulphide filter cake recovered for further processing if required. The filtrate from the bleed filter is recycled to the bleed final precipitator. Wash solution from the tails filters enters 200-V8 on an intermittent basis as noted in Section 4.3.1 above.

The overflow from 200-V8 is pumped through clarifier 200-F4, and is directed to the liquid waste disposal system drums.

4.3.3 - Area 300 - Zinc Electrowinning (PFD-003)

The solution from the zinc electrowinning feed tank, 200-V5, is pumped through clarifier 300-F1 and enters the zinc solvent extraction mixer-settler, 300-V1. The solution is contacted with a stripping solution of 60 percent TBP/40 percent Solvesso 150 flowing countercurrently in the solvent extraction mixer-settlers, 300-V1-V6. The lean solution overflows 300-V6 into the lean solution storage tank, 300-V23, from where it is forwarded to the copper recovery section, area 400.

The loaded solvent from 300-V1 enters the scrubbing mixer-settlers where it is scrubbed for iron removal (300-V7 and V8) and copper removal (300-V9), using strong electrolyte. The loaded scrub solution is mixed with the zinc extraction feed in 300-V1. The scrubbed solvent enters the solvent strippers, 300-V10-V15, where it is contacted countercurrently with spent electrolyte. From 300-V15, the lean solvent flows into the solvent storage tank, 300-V22, from where it is recirculated to the solvent extraction mixer-settler, 300-V6.

The loaded electrolyte overflows the zinc solvent stripper, 300-V10, into the purification feed tank, 300-V24. The electrolyte

is then pumped through the carbon column, 300-V16, for trace organics removal and enters the zinc purification vessels, 300-V17-19, where it is mixed with zinc oxide and zinc powder. The slurry from 300-V19 is pumped through the recessed-plate slurry filter, 300-F2. The filter cake is removed once every 24 hours and is discarded. Consideration will be given to returning the filter cake to the leach. The filtrate enters the electrowinning feed tank, 300-V20, where it is mixed with hydrochloric acid in order to control the pH at 1.0. The electrolyte is then pumped to the electrowinning cells 300-X1A-F. A portion of the strong electrolyte is directed to the brine addition tank, 300-V25, where it is mixed with sodium chloride and then used as scrub solution in the zinc solvent scrubbers.

The electrowinning cells are of the diaphragm type with the cathodes being continuously subjected to vigorous air agitation. Chlorine gas and spent anolyte are withdrawn from the anode compartments and enter the stripped liquor tank, 300-V21. The tank is agitated with air and the air/chlorine mixture is drawn through the demister to the chlorinators in area 100. Stripped liquor (spent electrolyte) is then recirculated to the zinc solvent strippers via stripped liquor pump, 300-P3.

Items 300-V1-V8 and 300-V22 are nitrogen blanketted to prevent oxidation of the ferrous ion.

4.3.4 - Area 400 - Copper Recovery (PFD-400)

The lean solution from zinc electrowinning passes through carbon column 400-V17 and enters the copper solvent extraction mixer-settlers, 400-V1 and V2, where it is contacted countercurrently with the 90 percent Kerosene/10 percent Acorga P5300 solvent. The lean solution overflows 400-V2 into the lean solution storage tank from where it is recycled to the 2nd stage leach. The extraction mixer-settlers, 400-V1 and V2, are air agitated for oxidation of cuprous ion to cupric.

The loaded solvent enters the copper solvent scrubbers, 400-V3 and V4, where it is scrubbed countercurrently with water pumped from the regen/scrub feed storage tank, 400-V8. The loaded scrub solution is directed to the effluent treatment system and is also used as filter wash water for the tails filter in area 100. The filter wash water is stored in the filter wash tank, 400-V16.

The scrubbed solvent flows into the copper solvent strippers, 400-V5 and V6, where it is stripped using filtrate from the copper sulphate filter, 400-F1.

The stripped solvent enters the copper solvent regenerator, 400-V7, where it is contacted with water from the regen/scrub feed tank, 400-V8. The regenerated solvent overflows into the solvent storage tank, 400-V12 and is then recirculated to the SX mixer-settler, 400-V2.

Pregnant solution from the copper solvent stripper, 400-V5, overflows into the pregnant solution surge tank, 400-V9. The solution is pumped through the carbon column, 400-V15, for trace organics removal, into mixing tank 400-V14. Sulphuric acid from a drum is added via metering pump 400-P6 in order to obtain an acid concentration of 170 g/l for copper sulphate crystallization. From 400-V14, the solution is pumped into the crystallizer, 400-X1, where it is cooled on a batchwise basis to 10°C. The copper sulphate slurry is pumped through the recessed-plate copper sulphate filter, 400-F1. The filter cake is water-washed every 12 hours and is recovered for sale. Wash water effluent is used as make-up to the regen/scrub feed tank, 400-V8.

Filtrate enters stripped liquor tank, 400-V11, where it is heated to 25°C and then circulated, after water makeup, to the copper solvent stripper, 400-V6.

4.3.5 - Area 500 - Lead Recovery and Off-Gas Scrubbing (PFD-005)

Lead chloride crystals from area 200 are dried in the lead chloride dryer, 500-DR1 on a batchwise basis. The dried crystals are manually dumped into a closed-top transportation vessel, 500-V3A or B, and sent to the crusher, 500-X2. The crushed solids are returned, using the transportation vessel and the vessel is used as the feed tank to the cell. The other lead chloride vessel is sent to the drier to await the next batch. The lead is manually added on a periodic basis to the electrolytic cell, 500-X1. The cell is of the fused-salt type with horizontal monopolar graphite anodes operating at a temperature of 450°C. Molten lead metal is siphoned into a mould by means of a lead vacuum chamber.

Chlorine is drawn from the cell due to the negative pressure developed by chlorine scrubber blower, 500-C1, and diluted with air. The gas is then mixed with chlorine off-gas from the chlorinators in area 100 and passes through the chlorine scrubber, 500-V1. The chlorine scrubber blower, 500-C1, vents the scrubbed gases to atmosphere. The chlorine scrubber utilizes a 10 percent solution of caustic as the scrubbing solution. The caustic is pumped from the scrubber feed tank, 500-V4 and recycled through the scrubber. The tank is fed by 50 percent caustic solution from the caustic drum.

The off-gases from the dryer and fumes from the lead chloride hopper together with off-gases from areas 100 and 200, are drawn by means of the off-gas blower, 500-C2, through the cyclonic off-gas scrubber, 500-V2, where the gases are water-scrubbed in order to remove particulates and any soluble gases present. The scrubbed gas is discharged to atmosphere.

The wash water for all the filters except the tailings units is stored in the filter wash water tank, 500-V5, and is pumped to the filters by filter wash water pump, 500-P3.

4.4 - Materials of Construction

Ferric chloride solution is an extremely corrosive fluid especially at high temperatures. There are basically three temperature levels in the pilot plant and the materials of construction used at each level require careful selection. Materials have been specified, on the whole, to minimize capital cost, although more expensive materials have been included at certain points in the plant in order to evaluate their suitability should the plant eventually be scaled up to a commercial size.

4.4.1 - High Temperature (100-108°C)

The temperature of both leach stages has to be maintained as high as possible to obtain the highest efficiency of extraction. The 1st and 2nd stage leach vessels have thus been specified to withstand a temperature of 108°C.

The 1st stage leach vessels could be constructed from three different materials, the 2nd stage leach vessels are all of FRP.

(a) Titanium - Titanium has been selected for the first of the 1st stage leach vessels in order to obtain data as to its suitability for a high temperature environment. Although the material is considerably more expensive than those used elsewhere in the leach section, it is thought worthwhile to include it as part of the test regime.

(b) FRP - Fibre-reinforced plastic has been specified as the material of construction for the second vessel in the 1st stage leach as well as for all three 2nd stage leach vessels. The material used in these vessels is "Hetron 197P", manufactured by Ashland Chemicals. The material is guaranteed to a temperature of 250°F (120°C) which is more than adequate for the pilot plant operation. FRP is considerably cheaper than titanium or lined vessels on this scale of operation.

(c) Glass Lined Carbon Steel - In commercial sizes it may be economical to use lined vessels instead of FRP or titanium. Ontario Research Foundation has a glass-lined vessel which could be used for the 3rd vessel in the 1st stage leach, in order to test its suitability in this medium.

A second glass-lined vessel has been specified as a cooling vessel in order that the temperature of the 2nd stage leach pulp is reduced to 85°C before pumping through the tails filter. Cooling of the pulp in this way considerably reduces the cost of the tails filters (see Section 4.4.2).

The method of heating the vessels has been purposely varied in order to obtain data on the various methods utilized.

- A titanium heating element is utilized on the first vessel in the 1st stage leach.

- An electrical heat tracing system is wrapped around each of the FRP vessels. The coils are then insulated and covered with an FRP skin for protection.

- The glass-lined vessels are each provided with a jacket. The third vessel in the 1st stage leach is heated by utilizing low pressure steam in the jacket whereas the 2nd stage leach cooling vessel is cooled by passing cooling water through the jacket.

All vessels are equipped with titanium agitators.

Both the 1st stage slurry pump and the tails filter pump are double-diaphragm metering pumps. The hot ferric chloride solution is pumped through a teflon diaphragm. Ball valves and seats are made of Hastelloy C.

All piping is FRP.

4.4.2 - Medium Temperature (80-100°C)

The most economical material for all vessels and piping in this temperature range is FRP. FRP has also been chosen for the chlorinators and storage tanks. Tower packing is polypropylene.

Pumps are double diaphragm metering pumps using the same materials as in Section 4.4.1.

Filters are of the plate-and-frame type. Polypropylene plates are used with polyester filter cloths. Polypropylene is acceptable to approximately 85-90°C, above which Kynar is required, which increases the price of the filter by a factor of 3.

Agitators are FRP coated carbon steel.

4.4.3 - Low Temperature (10°C-Ambient)

All vessels operating at the low temperature level will be fabricated from either FRP or PVC. Piping is flexible PVC tubing.

Diaphragm pumps have PVC ends with teflon diaphragms. Low head pumps are either squeegie or peristaltic with hypalon liners or PVC tubing.

Filters have polypropylene plates with polypropylene filter cloths.

Agitators are FRP coated carbon steel.

Blowers are fabricated in plastic.

Mixer/settlers are fabricated in perspex.

4.5 - Equipment List

4.5.1 - Area 100 - Leaching Circuit

Item	No. Required	Description
100-AG1-AG3	3	1st Stage Leach Agitators T - Propeller M - Titanium D - Electric Motor (TEFC)
100-AG4-AG7	3	2nd Stage Leach Agitators Identical to 100-AG1 - AG3
100-AG8A/B		Salt Solution Mixers T - Propeller M - C.S./Coated D - Electric Motor (TEFC)
100-F1	1	Pregnant Solution Clarifier T - Cartridge C - 66 L/h @ 500 ppm solids M - Glass-Lined Carbon Steel
100-F2 A/B	2	Tails Filters T - Recessed Plate Filter Press C - 74 L/h solution; 5.65 kg/h solids S - 600 mm x 600 mm x 17 chamber M - Polypropylene Plates; Polyester Filter Cloth
100-H1	1	Concentrate Feeder Package T - Tubular Spiral Conveyor Package C - 10 kg/h solids (S.G.5) D - Electric Motor (TEFC) R - Hopper sized for 7 days capacity
100-P1	1	1st Stage Slurry Pump T - Double Diaphragm C - 78 L/h, 180 kPa (gauge) discharge M - Teflon tube D - Electric Motor (TEFC)
100-P2	1	Pregnant Solution Pump T - Double Diaphragm C - 66 L/h, 345 kPa (g) discharge M - Teflon tube D - Electric Motor (TEFC)

Item	No. Required	Description
100-P3	1	Tails Filter Pump T - Double Diaphragm C - 74 L/h, 690 kPa (g) discharge M - Teflon Tube D - Electric Motor (TEFC)
100-P4	1	1st Stage Leach Recycle Pump T - Double Diaphragm C - 78 L/h, 180 kPa (g) discharge M - Teflon Tube D - Electric Motor (TEFC)
100-P5	1	2nd Stage Leach Recycle Pump T - Double Diaphragm C - 59.4 L/h, 180 kPa (g) discharge M - Teflon Tube D - Electric Motor (TEFC)
100-P6	1	HCl Metering Pump T - Diaphragm Metering (3 head) C - 2 x 2.6 L/h : 1 x 1.0 L/h M - PVC/Teflon D - Electric Motor (TEFC)
100-P7	1	Salt Solution Pump T - Diaphragm Metering C - 7.5 L/h, 60 kPa (g) discharge M - Polyethylene/Teflon D - Electric Motor (TEFC)
100-RV1	1	Rotary Valve C - 10 kg/h concentrate M - Plastic D - Electric Motor (TEFC)
100-V1	1	1st Stage Leach Vessel T - Vertical Cylindrical C - 170 litres S - 600 mm Ø x 600 mm t/t M - Titanium R - c/w Ti heating coil
100-V2	1	1st Stage Leach Vessel T - Vertical Cylindrical C - 170 litres S - 600 mm Ø x 600 mm t/t M - FRP R - c/w external electrical heating coil

Item	No. Required	Description
100-V3	1	1st Stage Leach Vessel T - Vertical Cylindrical C - 170 litres S - 600 mm Ø x 600 mm t/t M - Glass lined C.S. R - c/w steam jacket
100-V4-6	3	2nd Stage Leach Vessel Indentical to 100-V2
100-V7	1	2nd Stage Leach Cooling Vessel Identical to 100-V3 except jacket is used for cooling
100-V8	1	Pregnant Solution Tank T - Vertical Cylindrical C - 660 litres S - 750 mm Ø x 1500 mm ht M - FRP
100-V10	1	2nd Stage Chlorinator T - Packed Column with Storage Tank S - (column) - 300 mm Ø x 3,000 mm ht. M - FRP Packing - Polypropylene Saddles C (tank) - 950 litres S (tank) - 1,000 mm Ø x 1,200 mm ht. M - FRP R - c/w electrical heating coil
100-V11	1	1st Stage Chlorinator T - Packed Column with Storage Tank S (column)) identical to M (column)) 100-V9 Packing) C (tank) - 1,020 litres S (tank) - 1,000 mm Ø x 1,300 mm ht. M - FRP R - c/w electrical heating coil
100-V12A/B	2	Salt Solution Mixing Tank T - Vertical Cylindrical C - 98 litres S - 500 mm Ø x 500 mm ht. C - FRP/PVC
100-V13	1	Settling Cone T - Conical-Bottomed Tank C - 280 litres S - 500 mm Ø x 1,500 mm straight side M - FRP

4.5.2 - Area 200 - Silver Recovery and Bleed Treatment

Item	No. Required	Description
200-AG1-AG3	3	Silver Cementation Agitators T - Propeller M - CS/Coated D - Electric Motor (TEFC)
200-AG4	1	Bleed Oxidizer Agitator T - Propeller M - CS/Coated D - Electric Motor (TEFC)
200-AG5	1	Bleed Precipitator Agitator Identical to 200-AG4
200-F1	1	Cement Filter T - Plate and Frame Filter Press C - 66 L/h @ 35 g/h solids S - 300 mm x 300 mm x 6 chamber M - Polypropylene Plates; Polyester Backing
200-F2	1	Lead Chloride Filter T - Recessed Plate Filter Press C - 79.2 L/h @ 0.81 kg/h solids S - 450 mm x 450 mm x 9 chamber M - Polypropylene Plates/Filter Cloth
200-F3	1	Bleed Filter T - Recessed Plate Filter Press C - 77.2 L/h @ 12 kg/h solids S - 600 mm x 600 mm x 11 chamber M - Polypropylene Plates/Filter Cloth
200-F4	1	Bleed Clarifier T - Cartridge C - 77.2 L/h @ 500 ppm Solids M - Glass-Lined Carbon Steel
200-H1	1	Copper Powder Feeder Package T - Volumetric Feeder Package C - 129 g/h M - C.S. R - Hopper capacity 21 kg D - Electric Motor

Item	No. Required	Description
200-P1	1	Cementation Filter Pump T - Double Diaphragm C - 66 L/h, 690 kPa (g) discharge M - Teflon Tube D - Electric Motor (TEFC)
200-P2	1	Lead Chloride Filter Pump T - Diaphragm C - 80 L/h, 690 kPa (g) discharge M - PVC/Teflon D - Electric Motor (TEFC)
200-P3	1	Zinc Extraction Feed Pump T - Diaphragm C - 66 L/h, 345 kPa (g) discharge M - PVC/Teflon D - Electric Motor (TEFC)
200-P4	1	Oxidizer Discharge Pump T - Diaphragm C - 78 L/h, 690 kPa (g) discharge M - PVC/Teflon D - Electric Motor (TEFC)
200-P5	1	Bleed Filter Recycle Pump Identical to 200-P4
200-P6	1	Bleed Discharge Pump Identical to 200-P4
200-P7	1	Tails Filter Wash Pump T - Diaphragm C - 60 L/h, 690 kPa (g) discharge M - PVC/Teflon D - Electric Motor (TEFC)
200-RV1	1	Rotary Valve C - 125 g/h copper powder M - Plastic D - Electric Motor (TEFC)
200-V1-V3	3	Silver Cementation Mixing Vessels T - Vertical Cylindrical C - 42 litres S - 300 mm Ø x 600 mm t/t M - FRP R - c/w electrical heating coil
200-V4A/B	2	Lead Chloride Crystallization Vessels See 200-X1A/B

Item	No. Required	Description
200-V5	1	Zinc Extraction Feed Tank T - Vertical Cylindrical C - 1,420 litres S - 1,000 mm Ø x 1,800 mm ht. M - FRP/PVC
200-V6	1	Copper Stripped Solution Carbon Column T - Vertical Tube Packing - Activated Carbon S - 150 mm Ø x 1,250 mm ht. M - Glass/Perspex
200-V7	1	Bleed Oxidizer T - Conical Bottom Tank C - 530 litres S - 750 mm Ø x 1,200 mm ht. (flat side) M - FRP R - c/w agitator and PVC sparger
200-V8	1	Bleed Final Precipitator T - Conical Bottom Tank C - 530 litres S - 750 mm Ø x 1,200 mm ht. (flat side) M - FRP
200-V9	1	Tails Filter Wash Tank T - Vertical Cylindrical C - 160 litres S - 500 mm Ø x 800 mm ht. M - FRP/PVC
200-X1A/B	2	Lead Chloride Crystallizer Package T - Storage tank c/w pump, chiller C - 800 litres storage M - FRP R - To cool from 85 to 15°C over 12-hr period

4.5.3 - Area 300 - Zinc Electrowinning

Item	No. Required	Description
300-AG1-AG6	6	Zinc SX Pump Mixers T - Pump-Mixer M - CS/Coated D - Air Motor
300-AG7-AG9	3	Zinc Solvent Scrubber Pump Mixers Identical to 300-AG1-A66
30U-AG10-AG15	6	Zinc Solvent Stripper Pump Mixers Identical to 300-AG1-AG6
30U-AG16-AG18	3	Zinc Purification Agitators T - Propeller M - CS/Coated D - Electric Motor (TEFC)
300-AG19	1	Electrowinning Feed Agitator Identical to AG16-AG18
300-AG20	1	Brine Agitator T - Propeller M - CS/Coated D - Electric Motor (TEFC)
300-F1	1	Solution Clarifier T - Cartridge Filter C - 66 L/h @ 500 ppm Solids M - Glass-Lined Carbon Steel
300-F2	1	Zinc Electrolyte Filter T - Recessed Plate Filter Press C - 232 l/h @ 2 kg/h Solids S - 600 mm x 600 mm x 6 Chamber M - Polypropylene Plates and Cloths
300-P1	1	Zinc Filter Pump T - Diaphragm C - 232 L/h, 690 kPa (g) discharge M - PVC/teflon D - Electric Motor (TEFC)
300-P2	1	Electrowinning Feed Pump T - Peristaltic/Squeegie C - 115 L/h, 60 kPa (g) discharge M - Plastic D - Electric Motor (TEFC)

Item	No. Required	Description
300-P3	1	Stripped Liquor Pump T - Peristaltic/Squeegie C - 115 L/h, 60 kPa (g) discharge M - Plastic D - Electric Motor (TEFC)
300-P4	1	Purification Feed Pump T - Peristaltic/Squeegie C - 115 L/h, 60 kPa (g) discharge M - Plastic D - Electric Motor (EX PROOF)
300-P5	1	Solvent Recirculation Pump T - Peristaltic/Squeegie C - 230 L/h, 60 kPa (g) discharge M - Plastic D - Electric Motor (EX PROOF)
300-P6	1	Lean Solution Transfer Pump T - Peristaltic/Squeegie C - 74.4 L/h, 60 kPa (g) discharge M - Plastic D - Electric Motor (EX PROOF)
300-P7	1	Scrub Solution Pump T - Peristaltic/Squeegie C - 6 L/h, 60 kPa (g) discharge M - Plastic D - Electric Motor (TEFC)
300-V1-V6	6	Zinc SX Mixer Settlers T - Mixer-Settler C - 74 L/h aqueous; 230 L/h organic S - dims (mm) L : W : H Mixer 225 225 225 Settler 600 150 500 M - Perspex
300-V7-V9	3	Zinc Solvent Scrubbers T - Mixer-Settler C - 6 L/h aqueous (+ 70 L/h recycle); 230 L/h Organic S - dims (mm) L : W : H Mixer 185 185 185 Settler 600 150 500 M - Perspex

Item	No. Required	Description
300-V10-V15	6	Zinc Solvent Strippers T - Mixer-Settler C - 116 L/h aqueous; 230 L/h organic (+ 60 L/h recycle) S - dims (mm) L : W : H Mixer 250 250 250 Settler 625 175 500 M - Perspex
300-V16	1	Carbon Column T - Vertical Tube Packing - Activated Carbon S - 150 mm Ø x 1,250 mm ht. M - Glass/Perspex
300-V17-V19	3	Zinc Purification Vessels T - Vertical Cylindrical C - 530 litres S - 750 mm Ø x 1,200 mm ht. M - FRP/PVC
300-V20	1	Electrowinning Feed Surge Tank T - Vertical Cylindrical C - 530 litres S - 750 mm Ø x 1,200 mm ht. M - FRP/PVC
300-V21	1	Stripped Liquor Tank T - Vertical Cylindrical C - 800 litres (liquid) S - 1,000 mm Ø x 2,200 ht. M - FRP/PVC R - c/w demister
300-V22	1	Solvent Storage Tank T - Vertical Cylindrical C - 530 litres S - 750 mm Ø x 1,200 litres M - FRP/PVC
300-V23	1	Lean Solution Storage Tank T - Vertical Cylindrical C - 940 litres S - 1,000 mm Ø x 1,200 mm ht. M - FRP/PVC
300-V24	1	Purification Feed Tank T - Vertical Cylindrical C - 940 litres S - 1,000 mm Ø x 1,200 mm ht. M - FRP/PVC

Item	No. Required	Description
300-V25	1	Brine Addition Tank T - Vertical Cylindrical C - 78 litres S - 500 mm Ø x 1,000 mm ht. M - FRP/PVC
300-X1A-F	6	Electrowinning Cells T - Diaphragm Cell C - 3.52 kg/h Zn metal S - 2,800 amps, 7 m^2 total Cathode Area M - Polycarbonate Cell, Dynel Cloth Diaphragms Aluminum Cathodes DSA Anodes R - c/w (i) air spargers (ii) rectifier

4.5.4 - Area 400 - Copper Recovery

Item	No. Required	Description
400-F1	1	Copper Sulphate Filter T - Recessed Plate Filter Press C - 35 L/h @ 4% Solids S - 450 mm x 450 mm x 12 chamber M - Polypropylene Plates/Cloths
400-AG1/ AG2	2	Copper SX Pump Mixers T - Pump-Mixer M - CS/Coated D - Air Motor
400-AG3 AG4	2	Copper Solvent Scrubber Pump Mixers Identical to 400-AG1/AG2
400-AG5/ AG6	2	Copper Solvent Stripper Pump Mixers Identical to 400-AG1/AG2
400-AG7	1	Copper Solvent Regenerator Pump Mixers Identical to 400-AG1/AG2
400-AG8	1	Mixing Tank Agitator T - Propeller M - C.S./Coated D - Electric Motor (TEFC)
400-AG9	1	Stripped Liquor Tank Agitator T - Propeller M - C.S./Coated D - Electric Motor (TEFC)
400-P1	1	Lean Solution Transfer Pump T - Diaphragm C - 74.4 L/h, 180 kPa (g) discharge M - PVC/Teflon D - Electric Motor (EX PROOF)
400-P2	1	Regen/Scrub Feed Pump T - Peristaltic/Squeegie C - 35 L/h, 60 kPa (g) discharge M - Plastic D - Electric Motor (EX PROOF)

Item	No. Required	Description
400-P3	1	Pregnant Solution Transfer Pump T - Peristaltic/Squeegie C - 16.2 L/h, 60 kPa (g) discharge M - Plastic D - Electric Motor (EX PROOF)
400-P4	1	Crystallizer Feed Pump T - Peristaltic/Squeegie C - 35 L/h, 60 kPa (g) discharge M - Plastic D - Electric Motor (TEFC)
400-P5	1	Stripped Liquor Pump T - Peristaltic/Squeegie C - 16.2 L/h, 60 kPa (g) discharge M - Plastic D - Electric Motor (TEFC)
400-P6	1	Sulphuric Acid Metering Pump T - Diaphragm Metering C - 1 L/h M - C.I./Teflon D - Electric Motor (TEFC)
400-P7	1	Filter Feed Pump T - Diaphragm C - 36 L/h, 690 kPa (g) discharge M - Alloy 20/Teflon D - Electric Motor (TEFC)
400-P8	1	Filter Wash Pump T - Diaphragm C - 35 L/h, 690 kPa (g) discharge M - PVC/Teflon D - Electric Motor (EX PROOF)
400-P9	1	Solvent Recirculation Pump T - Peristaltic/Squeegie C - 81 L/h, 60 kPa (g) discharge M - Plastic D - Electric Motor (EX PROOF)
400-V1-V2	2	Copper SX Mixer-Settlers T - Mixer-Settler C - 74.4 L/h aqueous; 81 L/h organic S - dims (mm) L : W : H Mixer 255 255 255 Settler 450 110 500 M - Perspex R - c/w PVC air sparger

Item	No. Required	Description
400-V3-V4	2	Copper Solvent Scrubbers T - Mixer-Settler C - 20 L/h aqueous (+7 L/h recycle) 81 L/h organic S - dims (mm) L : W : H Mixer 135 135 135 Settler 400 100 500 M - Perspex
400-V5-V6	2	Copper Solvent Stripper T - Mixer-Settler C - 16.2 L/h aqueous; 81 L/h organic S - dims (mm) L : W : H Mixer 135 135 135 Settler 400 100 500 M - Perspex
400-V7	1	Copper Solvent Regenerator T - Mixer-Settler C - 15 L/h aqueous (+ 12 L/h recycle); 81 L/h organic S - Dims (mm) L : W : H Mixer 100 100 250 Settler 400 100 500 M - Perspex
400-V8	1	Regen/Scrub Storage Tank T - Vertical Cylindrical C - 440 litres S - 750 mm Ø x 1,000 mm ht. M - FRP/PVC
400-V9	1	Pregnant Solution Surge Tank T - Vertical Cylindrical C - 210 litres S - 600 mm Ø x 780 mm ht. M - FRP/PVC
400-V11	1	Stripped Liquor Tank T - Vertical Cylindrical C - 210 litres S - 600 mm Ø x 750 mm ht. M - FRP/PVC

Item	No. Required	Description
400-V12	1	Solvent Storage Tank T - Vertical Cylindrical C - 440 litres S - 750 mm Ø x 1,000 mm ht. M - FRP/PVC
400-V13	1	Lean Solution Storage Tank T - Vertical Cylindrical C - 940 litres S - 1,000 mm Ø x 1,200 mm ht. M - FRP/PVC
400-V14	1	Mixing Tank T - Vertical Cylindrical C - 210 litres S - 600 mm Ø x 750 mm ht. M - FRP/PVC
400-V15	1	Carbon Column T - Vertical Tube Packing - Activated Carbon S - 150 mm Ø x 1,250 mm ht. M - Glass/Perspex
400-V16	1	Filter Wash Tank T - Vertical Cylindrical C - 210 litres S - 600 mm Ø x 750 mm ht. M - FRP/PVC
400-V17	1	Carbon Column T - Vertical Tube Packing - Activated Carbon S - 150 mm Ø x 1,250 mm ht. M - Glass/Perspex
400-X1	1	Crystallizer Package T - Cooling Crystallizer C - 200 litres Storage S - Cool to 10°C (from ambient) in 2 hrs. R - Complete with tank, pump, heat exchanger

4.5.5 - Area 500 - Lead Recovery and Off-Gas Scrubbing

Item	No. Required	Description
500-AG1	1	Chlorine Scrubber Feed Tank Agitator T - Propeller M - CS/Coated D - Electric Motor (TEFC)
500-C1	1	Chlorine Scrubber Blower T - Centrifugal C - $5m^3/h$ M - Plastic D - Electric Motor (TEFC)
500-C2	1	Off-Gas Blower T - Centrifugal C - $25m^3/h$ M - Plastic D - Electric Motor (TEFC)
500-DR1	1	Lead Chloride Dryer T - Oven C - 200 Kg $PbCl_2$ (16% moisture) R - Operating Temp. 200°C (max) Positive Air Circulation Final Water Content 0.1%
500-P1	1	Chlorine Scrubber Feed Pump T - Peristaltic/Squeegie C - 90 L/h, 60 kPa (g) discharge M - Plastic D - Electric Motor (TEFC)
500-P2	1	Caustic Metering Pump T - Diaphragm Metering C - 2.5 L/h, 60 kPa (g) discharge M - PVC/Teflon D - Electric Motor (TEFC)
500-P3	1	Filter Wash Water Pump T - Diaphragm C - 50l/h at 690 kPa (g) discharge M - PVC/Teflon D - Electric Motor (TEFC)

Item	No. Required	Description
500-P4	1	Off-Gas Scrubber Recirculation Pump T - Peristaltic/Squeegie C - 1200 L/h, 60 kPa (g) discharge M - Plastic D - Electric Motor (TEFC)
500-V1	1	Chlorine Scrubber T - Packed Column P - Polypropylene Saddles S - 300 mm Ø x 3,000 mm ht. M - FRP
500-V2	1	Off-Gas Scrubber T - Cyclonic with Separator S - Scrubber - 180 mm Ø x 600 mm ht. Separator - 600 mm Ø x 800 mm ht. M - FRP
500-V3A/B	2	Lead Chloride Vessels T - Vertical Cylindrical C - 75 litres S - 400 mm Ø x 600 mm ht. M - PVC/FRP R - c/w Wheels
500-V4	1	Scrubber Feed Tank T - Vertical Cylindrical C - 200 litres S - 600 mm Ø x 800 mm ht. M - PVC
500-V5	1	Filter Wash Tank T - Vertical Cylindrical C - 75 litres S - 400 mm Ø x 600 mm ht. M - PVC
500-X1	1	Lead Chloride Electrolytic Cell Package T - Fused Salt Cell with Horizontal Monopolar Anodes C - 1.1 kg/h lead production S - 300 A @ 450°C Anodes (2) 90 mm x 250 mm Cathode (1) 200 mm x 250 mm M - Electrodes - graphite Cell - Silica brick in steel shell

Item	No. Required	Description
		R - c/w (i) Vacuum chamber for lead removal (ii) Two graphite heating electrodes (iii) 500 A rectifier (iv) Rotary valve for lead chloride addition
500-X2	1	Lead Chloride Crusher T - Roll Crusher C - 200 Kg $PbCl_2$
500-X3	1	Chiller Package T - Refrigeration package for crystallization vessels

4.6 - Mass Balance - Major Streams

The compositions of the major streams in the pilot plant are tabulated below. The figures have been derived using available data and will require verification during plant operation.

The stream numbers are those appearing on the flowsheets.

4.6.1 - Feed and Products Distribution

Stream	1	2	3	4	5	6	7
Description	Concentrate	Lead Chloride Crystals	Silver/ Copper Cement	Zinc Product	Copper Sulphate Product	Tails	Bleed Solids
Total Flow kg/h (Dry Wt.)	10.0	1.01	33 (g)	3.10	0.66	6	2
Metals Content % (kg/h)							
Pb	8.75 (0.88)	74.5 (0.75)	-	-	-	0.9 (0.06)	5.2 (0.10)
Zn	28.1 (2.81)	-	-	100 (3.10)	-	0.1 (<0.01)	2.0 (0.04)
Cu	0.78 (0.08)	-	94 (31)	-	25.4 (0.17)	0.1 (<0.01)	0.2 (<0.01)
Ag g/t (g/h)	233 (2.33)	-	6 (2)	-	-	35 (21.00)	13.3 (0.03)
Fe	19.1 (1.91)	-	-	-	-	22.3(1.34)	26.1 (0.52)
Other	43.3 (4.33)	25.5 (0.26)	-	-	74.6 (0.49)	76.7(4.60)	66.5 (1.33)

4.6.2 - Major Internal Liquid Streams

Stream	8	9	10	11	12	13
Description	Pregnant Leach Solution	Solution to $PbCl_2$ Cryst.	Zinc SX Feed Solution	Copper SX Feed Solution	Leach Recycle Solution	Bleed Solution
Total Flow (L/h)	66.0	66.0	74.4	74.4	59.4	10.0
Metals Content G/L (kg/h)						
Pb	23.4 (1.55)	23.4 (1.55)	10.7 (0.79)	10.3 (0.77)	10.3 (0.61)	10.3 (0.10)
Zn	45.6 (3.01)	45.6 (3.01)	43.9 (3.27)	4.0 (0.30)	4.0 (0.24)	4.0 (0.04)
Cu	1.5 (0.10)	3.0 (0.20)	2.6 (0.20)	2.5 (0.19)	0.3 (0.02)	0.3 (0.003)
Ag mg/l (g/h)	34.6 (2.28)	3.0 (0.20)	2.7 (0.20)	2.7 (0.20)	2.7 (0.16)	2.7 (0.03)
Fe	59.1 (3.89)	59.1 (3.89)	52.3 (3.89)	52.2 (3.88)	52.2 (3.10)	52.2 (0.52)

4.7 - DRAWINGS

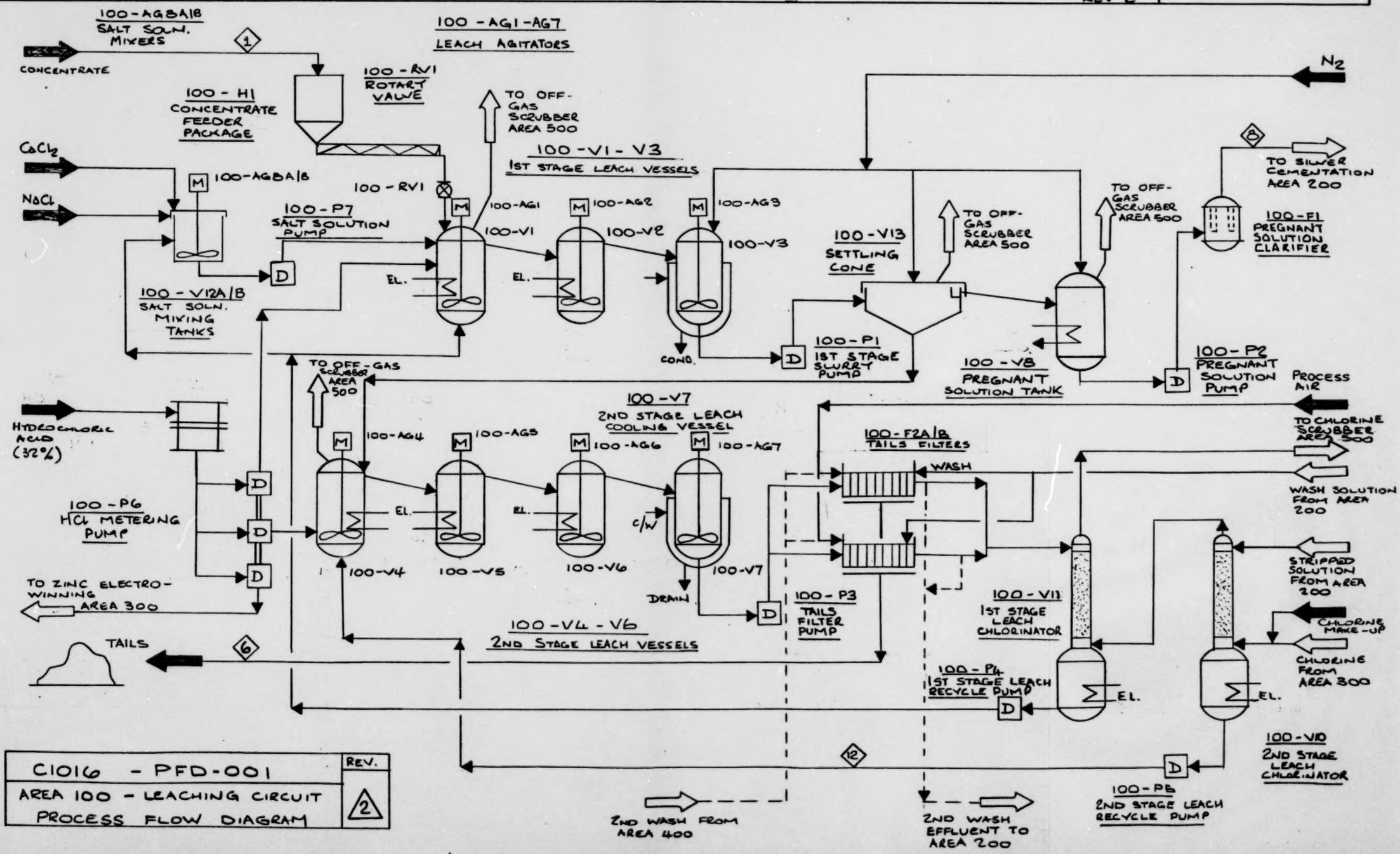
ADM Acres Davy McKee

SUBJECT: FERRIC CHLORIDE LEACH PILOT PLANT
PROCESS FLOW DIAGRAM

AREA 100
LEACHING CIRCUIT

REV 2

JOB NUMBER C1016
FILE NUMBER PFD - 001
SHEET 1 OF 1
BY DC DATE 14/2/84
APP _____ DATE _____



C1016 - PFD-001	REV.
AREA 100 - LEACHING CIRCUIT PROCESS FLOW DIAGRAM	2

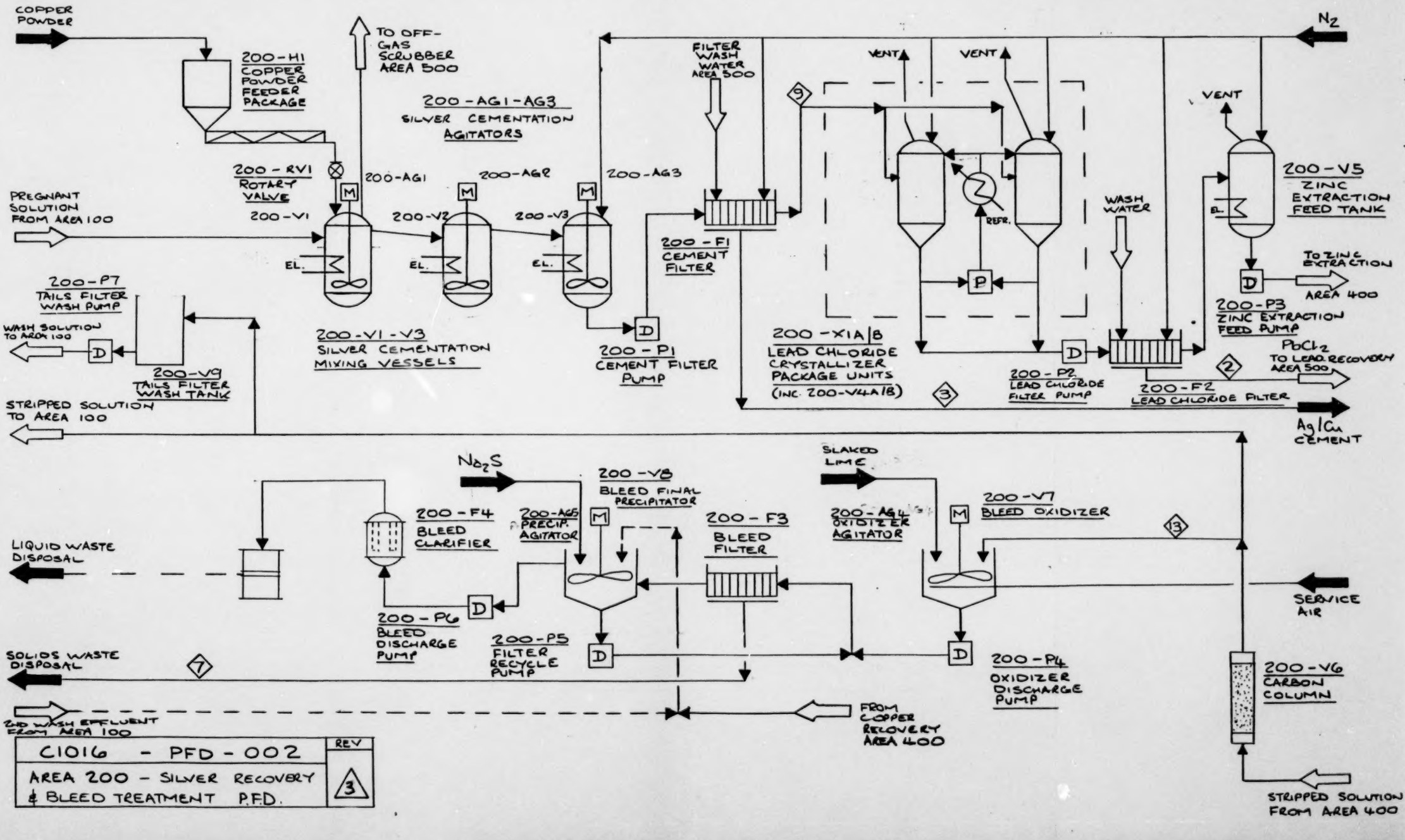
ADM Acres Davy McKee

SUBJECT:

DSS - FERRIC CHLORIDE LEACH PILOT PLANT
PROCESS FLOW DIAGRAM

AREA 200
SILVER & RECOVERY
BLEED TREATMENT
REV 3

JOB NUMBER	C1016
FILE NUMBER	PFD-002
SHEET	1 OF 1
BY	DC DATE 15/2/84
APP	DATE

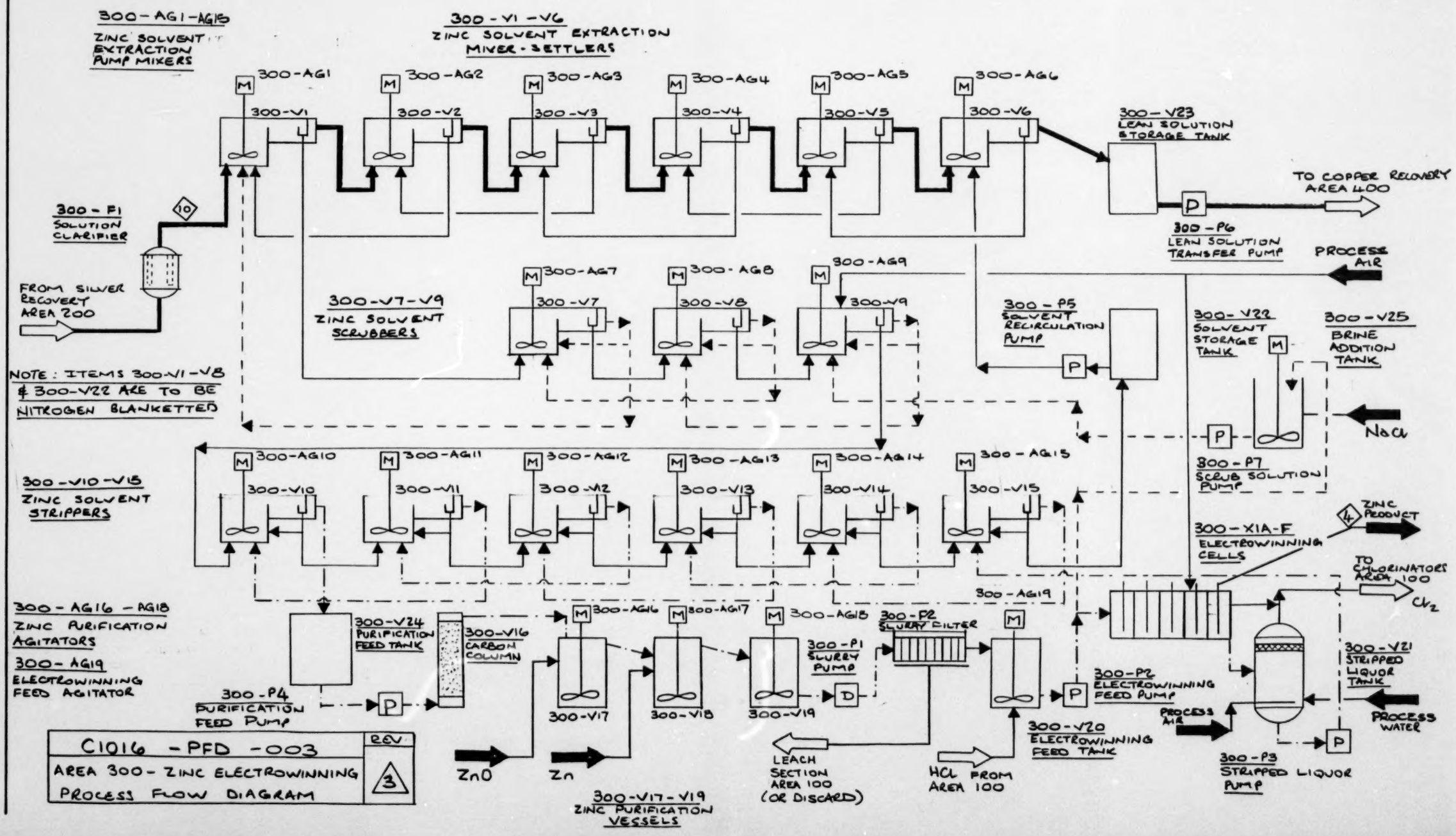


ADM Acres Davy McKee

SUBJECT: DSS - FERRIC CHLORIDE LEACH PILOT PLANT
PROCESS FLOW DIAGRAM

AREA 300
ZINC
ELECTROWINNING
REV 3

JOB NUMBER C1016
FILE NUMBER PFD - 003
SHEET 1 OF 1
BY DC DATE 16/2/94
APP _____ DATE _____



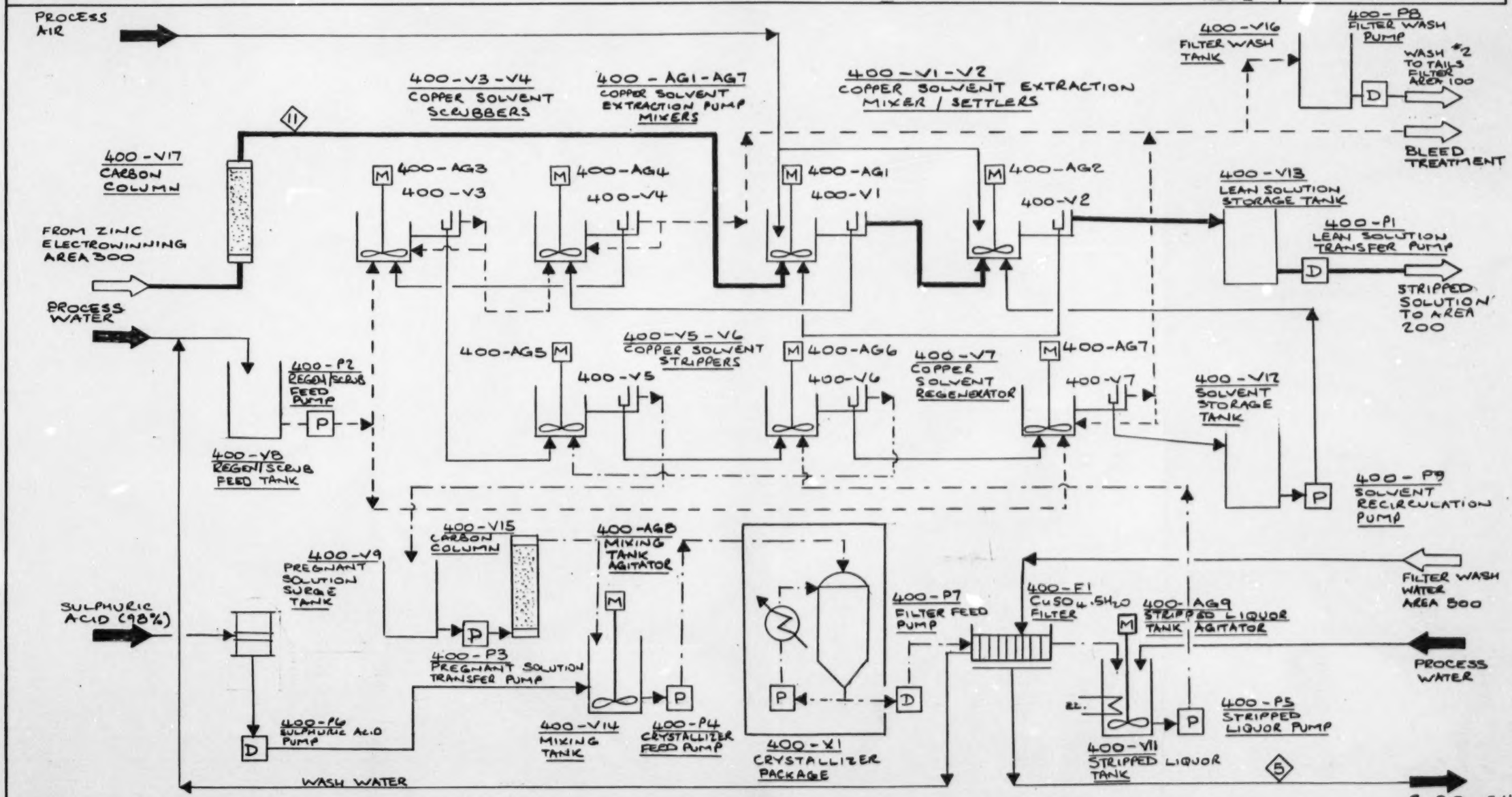
C1016 - PFD - 003	REV.
AREA 300 - ZINC ELECTROWINNING PROCESS FLOW DIAGRAM	3

ADM Acres Davy McKee

SUBJECT: DSS - FERRIC CHLORIDE LEACH PILOT PLANT
PROCESS FLOW DIAGRAM

AREA 400
COPPER
RECOVERY
REV 2

JOB NUMBER	C1016
FILE NUMBER	PFD - 004
SHEET	1 OF 1
BY	DC DATE 14/2/84
APP	DATE



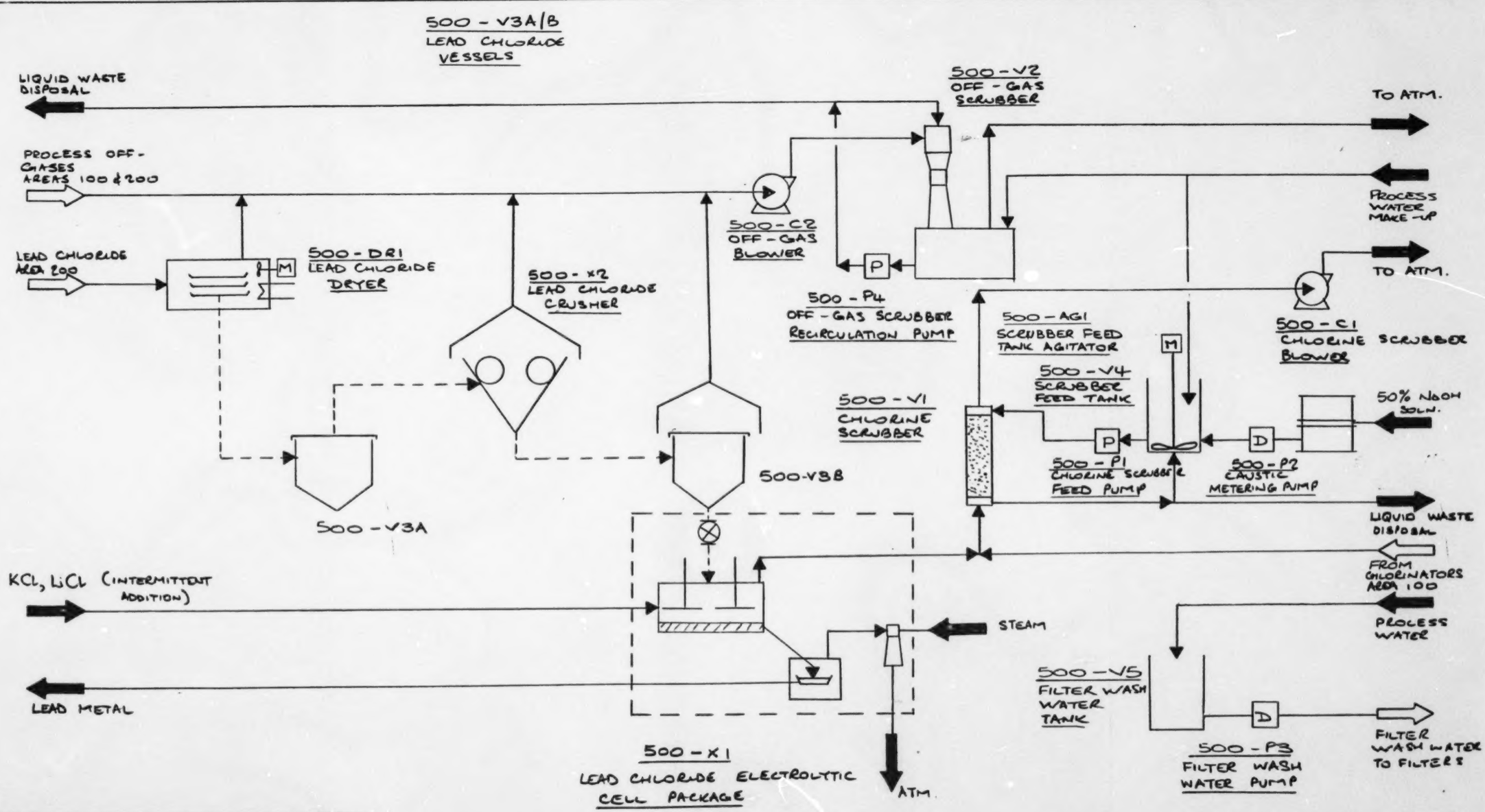
C1016 - PFD - 004	REV
AREA 400 - COPPER RECOVERY PROCESS FLOW DIAGRAM	2

ADM Acres Davy McKee

SUBJECT: DSS - FERRIC CHLORIDE LEACH PILOT PLANT
PROCESS FLOW DIAGRAM

AREA 500
LEAD RECOVERY & OFF-GAS SCRUBBING
REV 2

JOB NUMBER C1016
FILE NUMBER PFD-005
SHEET 1 OF 1
BY DC DATE 16/2/82
APP ____ DATE ____



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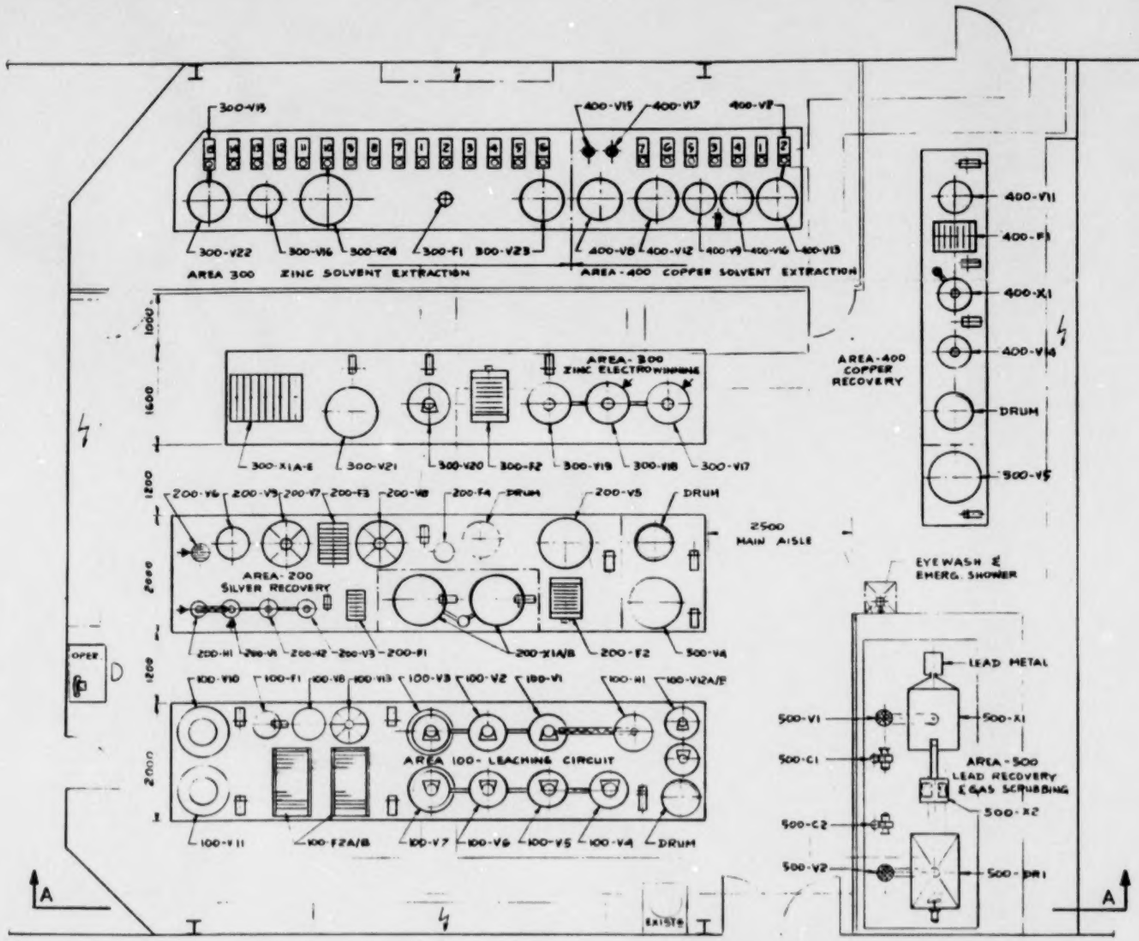
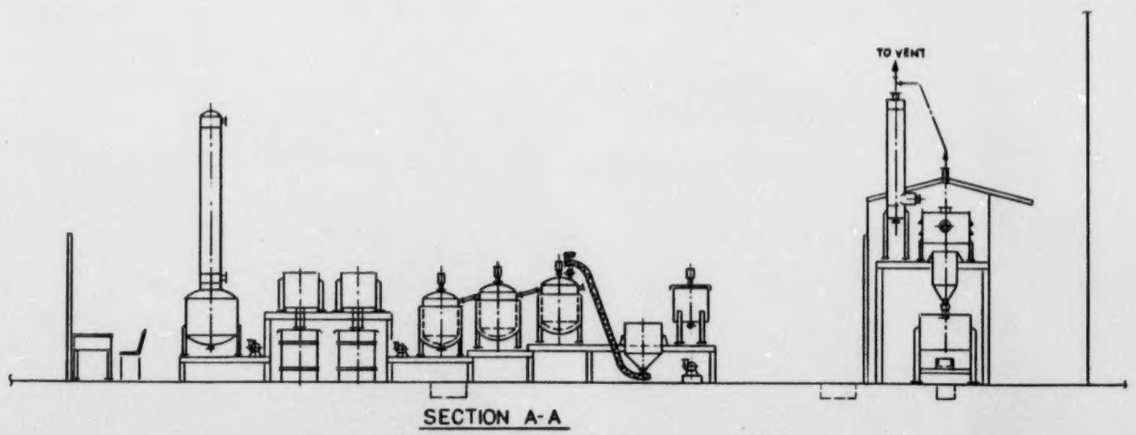
LEAD RECOVERY & GAS SCRUBBING - PFD	2





SECTION A-A

ADM DSS - CANMET

FERRIC CHLORIDE LEACH
PILOT PLANT
PRELIMINARY LAYOUT

C1016-M-001 | DATE MAR. 26/86 | SCALE 1:50 | A-23-001 | ACRES | PROJECT C-1016

REVISIONS

DRAWING No.

5 - PILOT PLANT COST ESTIMATE

5 - PILOT PLANT COST ESTIMATE

5.1 - Estimate Summary

5.1.1 - Introduction

The capital cost estimate for the ferric chloride leaching mini-plant has been produced from the equipment list, preliminary drawings and process description as presented in Section 4 of this report. This estimate is an order of magnitude type estimate and the overall accuracy is ±25 percent. All costs and prices are expressed in Canadian dollars as of April 1984.

5.1.2 - Capital Cost Estimate Summary

	Cost $Can
Major Equipment	368,000
Equipment Delivery	7,400
Major Equipment Erection	73,600
Civil/Structural	20,000
Building Modifications	11,800
Electrics	18,400
Piping	11,000
Instrumentation	14,700
Miscellaneous	10,000
Services	11,800
Organic Reagents	2,000
Construction Indirects	5,900
Engineering, Procurement and Management	110,400
Subtotal	665,000
Contingency 15%	99,800
Total	764,800
Cost saving due to use of existing ORF equipment	$ 65,800

5.2 - Basis of Estimate

(a) General

The capital cost estimate of the mini-plant has been produced based on the assumption that the plant will be installed at an existing testing facility, such as ORF using the flowsheet contained in this report. Consequently, no provision in the estimate has been made for a building, laboratory, offices, services, fire protection and liquid effluent disposal. Also the installation of the mini-plant has been assumed to be performed by employees of the testing facility.

(b) Major Equipment Costs

The cost of the major equipment has been producing using vendor budget type quotations for approximately 90 percent of the total equipment cost. The balance has been estimated using in-house experience. All the equipment prices are for new equipment. A separate cost saving from using some existing equipment at ORF has been shown in the estimate.

(c) Major Equipment Delivery and Erection

The costs of delivery and erecting the major equipment have been estimated by applying factors to the major equipment cost based on previous in-house experience on similar installations.

(d) Civil/Structural

The cost of civils and structures has been produced by applying a factor to the major equipment cost based on in-house experience. The amount of civil work is very limited. For equipment supports, both steel and wood are used depending on the duty.

(e) Building Modifications

An allowance has been made in the estimate for the cost of the building modifications, which is based on in-house experience. The building modifications are primarily to enclose the solvent extraction sections in a separate area.

(f) Electrics

An allowance has been made in the estimate for the cost of electrics, which is based on in-house experience. The costs mainly apply to the electrical connections to the electrolytic cells. The costs for the rectifiers for the electrolytic sections are included in the major equipment cost. Explosion-proof motors and electrical equipment are used in the solvent extraction areas.

(g) Piping

The cost of piping has been produced by applying a factor to the major equipment cost based on in-house experience. FRP piping is used for the higher temperature applications (>80°C) and flexible PVC hose is used elsewhere.

(h) Instrumentation

The cost of instrumentation has been produced by applying a factor to the major equipment cost based on in-house experience. The instrumentation consists primarily of indication of flows, temperatures, levels and pressures.

(i) Miscellaneous

An allowance has been made in the estimate for insulation and portable fire extinguishers.

(j) Services

An allowance has been made in the estimate for the provision of any additional services to those available at the testing facility.

(k) Reagents

An allowance has been made for the initial charge of solvents in the solvent extraction areas.

(l) Construction Indirects

The cost of construction indirects has been produced by applying a factor to the major equipment cost based on in-house experience. This cost covers the cost of construction supervision, equipment and tools, and office expenses.

(m) Engineering, Procurement and Management

The cost of engineering, procurement and management has been produced by applying a factor to the major equipment cost based on in-house experience.

Procurement costs are based on the assumption of single-source of speciality equipment items.

(n) Contingency

A contingency of 15 percent has been included to cover items that have been omitted or under-estimated because of the preliminary status of the mini-plant design. This is an integral part of the estimate and does not cover potential changes in scope, etc.

(o) Exclusions

The following items have been excluded from the capital cost estimate

- royalties and license fees
- supply of raw materials, reagents and services outside plant battery limits
- disposal of products, by-products and effluents outside plant battery limits
- purchase or lease cost of building
- financing and interest costs
- CANMET's costs
- escalation
- construction insurance
- laboratory and office facilities
- federal and provincial taxes.
- fire protection (sprinklers, hydrants etc.)

5.3 - Major Equipment Costs By Area

The following prices for equipment costs are based on new equipment. In most cases, the prices are based on quotations from manufacturers. Where quotations were not obtained the costs quoted are based on in-house or judgmental data.

Section 5.3.7 tabulates the cost savings due to usage of in-house equipment if the pilot plant were to be located at ORF's facilities.

5.3.1 - Area 100

	$ Can
Vessels	43,500
Agitators	22,000
Pumps	17,200
Filters	28,500
Feeders	6,600
Miscellaneous	2,000

Total for area 100 equipment = $119,800.

5.3.2 - Area 200

	$ Can
Vessels	9,200
Agitators	6,400
Pumps	12,100
Filters	26,200
Feeders	4,500
Crystallizers	5,000
Miscellaneous	2,000

Total for area 200 equipment = $65,400.

5.3.3 - Area 300

	$ Can
Vessels	19,500
Agitators	13,400
Pumps	8,700
Filters	14,900
Cells	23,000
Rectifier	6,000
Miscellaneous	2,000

Total for area 300 equipment = $87,500

5.3.4 - Area 400

	$ Can
Vessels	11,900
Agitators	4,900
Pumps	11,000
Filters	7,700
Crystallizer	3,700
Miscellaneous	2,000

Total for area 400 equipment = $41,200.

5.3.5 - Area 500

	$ Can
Vessels	11,900
Agitators	700
Pumps	2,400
Blowers	2,400
Dryer	10,000
Cells	9,000
Crusher	10,000
Miscellaneous	7,000

Total for area 500 equipment = $53,400.

5.3.6 - Totals - All Areas

		Equipment Cost ($ Can)
Area	100	119,800
Area	200	65,400
Area	300	87,500
Area	400	41,200
Area	500	53,400

Total equipment costs - all areas = $367,300

Say $368,000

5.3.7 - Equipment Available at ORF

	$ Can
Vessels	30,700
Agitators	4,800
Pumps	7,900
Blowers	2,400
Dryer	10,000
Crusher	10,000

Total cost of equipment available at ORF = $65,800.

5.4 - Schedule

A preliminary schedule for Phase II for the design, installation and operating of the mini-plant is presented below. This schedule reflects the use of the preliminary process design produced in Phase I and existing testing facilities such as ORF.

ADM Archer Daniels Midland

SCHEDULE SHEET

SHEET 1 OF 1

CHARGE NO. C1016.00

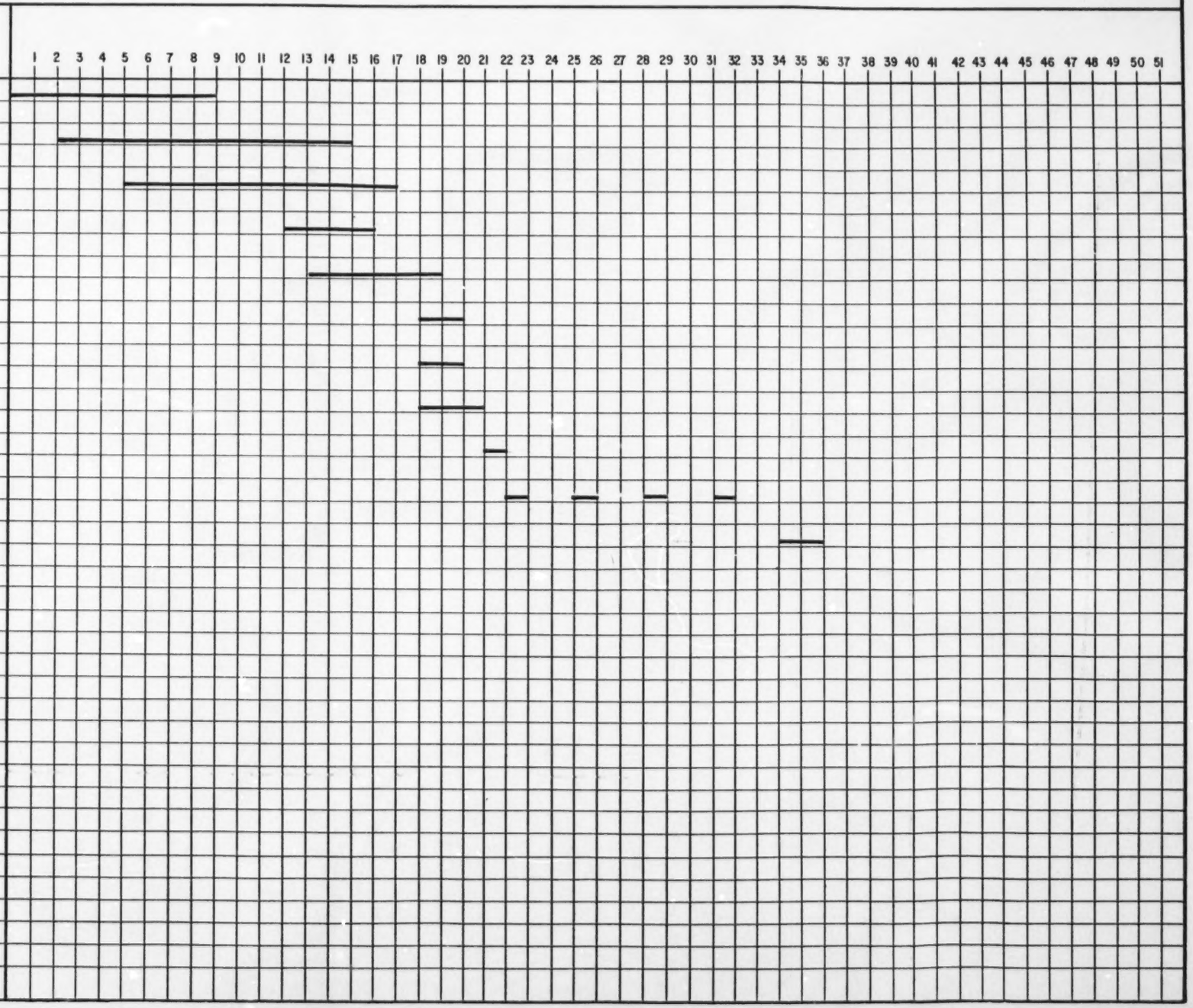
CLIENT DEPARTMENT OF SUPPLY AND SERVICES (CANMET)

PROJECT FERRIC CHLORIDE LEACHING MINI-PLANT

PREPARED BY _____ DATE March 23, 1984

PROJECT APPROVAL _____

WORK PACKAGE	DESCRIPTION
	PROCESS DESIGN
	DETAILED ENGINEERING
	EQUIPMENT PROCUREMENT
	CIVIL/STRUCTURAL INSTALLATION
	EQUIPMENT INSTALLATION
	ELECTRICAL INSTALLATION
	INSTRUMENTATION INSTALLATION
	PIPING INSTALLATION
	MINI-PLANT TESTING
	MINI-PLANT OPERATION
	MINI-PLANT DISMANTLING AND CLEAN-UP



Form 276A

APPENDIX I

APPENDIX I

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APPENDIX II

APPENDIX II

Patent Literature References

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4 292 281	8/81	Chilcote et al	Hanna
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4 346 062	8/82	Au	Occidental
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Canadian			
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Other			
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491 424	6/75	Agnew	--
498 645	1/77	Hurst	I.C.I.
498 645	1/77	Hurst	I.C.I.

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